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Investigation of Municipal Solid  
Waste and Paper to Alcohol Study

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# Investigation of Municipal Solid Waste to Alcohol Conversion for Army Use

by

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Rising costs of nonhazardous solid waste (SW) disposal, along with strict environmental regulations, limited landfill capacity, and the difficulty in siting new incinerators and landfills make SW disposal a costly, complex problem for Army installations. Converting SW to alcohol is one proposed way to lessen the SW disposal burden. This study evaluated and compared two different proposed processes for converting municipal solid waste (MSW) and waste paper to fuel ethanol and other byproducts. To convert the cellulosic fraction of MSW to fermentable sugars, the first process uses a single stage of dilute acid hydrolysis, and the second process uses enzymatic hydrolysis. Both processes promise to minimize the amount of waste material that would otherwise be landfilled or discarded. Ethanol—a byproduct of these processes—will continue to increase in value and market share as a fuel, fuel extender, or octane enhancer that can help reduce U.S. dependence on imported oil.

The study concluded neither process is presently ready for a single-step expansion to an economically viable, commercial-scale application. However, with continued development, both processes can serve as bases for detailed engineering and technology development studies leading to the design, construction, and operation of a commercial demonstration plant.

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## EXECUTIVE SUMMARY

### Background

Rising costs of nonhazardous solid waste (SW) disposal, along with strict environmental regulations, limited landfill capacity, and the difficulty in siting new incinerators and landfills make SW disposal a costly, complex problem for Army installations. Converting SW to alcohol is one proposed way to lessen the SW disposal burden. This report presents the results of a study that evaluated and compared the technical and economic feasibility of two processes for converting cellulosic materials (municipal solid waste [MSW] and waste paper) to fuel ethanol and other byproducts. The processes involve either dilute acid hydrolysis or enzymatic hydrolysis of the cellulosic fraction to produce fermentable sugars.

The first process, developed by the Tennessee Valley Authority (TVA), employs a single stage of dilute sulfuric acid hydrolysis to convert the cellulose and hemicellulose in MSW to fermentable sugars and furfural. Technical information was obtained from a proposal to process 400 to 1200 tons/day of MSW at a site-specific location in southern Indiana. The plant would be, in essence, an MSW-processing plant from which ethanol is the major byproduct, accounting for 31.4 percent of total projected revenues. Byproduct credits would also be taken for furfural, carbon dioxide, acetic acid, and electricity, as well as recyclable aluminum, ferrous metals, glass, and plastics recovered from the MSW.

The process proposed by the University of Alabama in Huntsville (UAH) employs an enzyme system (cellulase) to convert the cellulose in waste paper to fermentable sugars. The plant proposed by UAH would convert the waste paper generated daily from the Redstone Arsenal, in Alabama (16 to 27 ton/day) to ethanol. Byproducts from the UAH process include solid fuel, fusel oil, yeast, and carbon dioxide as well as the recyclables listed for the TVA process. An existing building at the Redstone Arsenal would be converted to house the processing equipment.

Both processes convert the sugars generated from cellulose hydrolysis to ethanol via conventional batch yeast (*Saccharomyces cerevisiae*) fermentation. The fermented beer is then upgraded to 190 proof ethanol by steam distillation. In the TVA process, the ethanol stream is processed through an anhydrous column to yield 199-proof ethanol, which is then denatured. In the UAH process, the ethanol stream is subjected to pervaporation, which selectively removes water from the azeotrope using a semipermeable membrane to yield 200 proof alcohol.

Each process responds to a specific need to minimize the amount of waste material to be landfilled or otherwise disposed of. Each process will help reduce U.S. dependence on imported liquid fuels by producing fuel ethanol as a byproduct. Ethanol, as a fuel, fuel extender, or octane enhancer, will continue to increase in value and market share as a way to reduce U.S. dependence on imported oil.

Both approaches to converting cellulosic materials to fermentable sugars have been the subject of much research. A literature review was conducted to ascertain the level of maturity of these two, and other, processes under study, and to determine the state-of-the-art technology in cellulose conversion to ethanol.

### Study

The primary source of information on the dilute acid hydrolysis process was a feasibility study prepared by TVA entitled *Preliminary Feasibility Study of the Recovery and Production of Recyclables, Chemicals, and Electricity From Municipal Solid Waste for Southern Indiana* (June 1990). The primary



source of information for the enzymatic hydrolysis process was a study prepared by UAH and Leo A. Daly, Inc. entitled *Ethanol Production Facility: Analysis/Feasibility Study* (September 1990).

These and other relevant documents were thoroughly reviewed from a technical and processing standpoint and the assumptions made for the economic evaluation were then assessed. Detailed study indicated that a direct economic comparison of the two processes for such disparate sizes could not be made; researchers therefore selected an MSW processing plant size of 400 tons/day and prepared scaled-up material balances based on each process. Where specific information was not explicitly available in the feasibility studies, reasonable engineering assumptions were made. Energy requirements were estimated and some calculation steps were also added to make the two processes comparable. For example, a front-end MSW classification section was incorporated into the enzymatic hydrolysis-based process because one was not included in the site-specific study. For simplicity, the 190-proof ethanol stream in both processes was dehydrated to neat ethanol by conventional extractive distillation.

Finally, preliminary evaluations were made on the economics of each generic plant within a common set of economic parameters. The sensitivities of the economics to changes in feedstock quality, plant size, tipping fee, return on investment, and other factors were determined. The process evaluations were not site-specific as were the feasibility studies. Researchers did not make any recommendations concerning the merits or disposition of the individual feasibility studies that were submitted.

## Results

Based on a careful review of the feasibility studies prepared by TVA and UAH, the published literature, information gathered by visits to TVA and UAH facilities, and an independent process scale-up and economic evaluation by the Institute of Gas Technology (IGT) the following observations were made:

- Use of MSW through recovery of recyclable materials and conversion of the cellulosic fraction to both useful and valuable byproducts can significantly reduce the burden on existing landfills for waste disposal.
- The production of fuel ethanol from MSW will help reduce U.S. dependence on imported liquid fuels.
- The scale-up of these processes from essentially batch, nonintegrated systems to large-scale commercial facilities in one step is not presently feasible. The degree of scale-up proposed by UAH is less ambitious than that by TVA. (Note that TVA enjoys significant industrial support in its process development efforts.)
- The acid hydrolysis stage of the TVA process has been tested in a 2 ton/day batch reactor. Fermentation of the hydrolyzate from the first stage has not yet been demonstrated. (Metals in the MSW may prove toxic to the yeast, among other considerations.)
- TVA laboratory and pilot plant data do not support the assumptions made in the process design for sugar yields from hydrolysis of WDF (waste-derived feedstock).
- The MSW sterilization stage of the UAH process has been tested in a 2 ton/day batch reactor (600 lb/batch). Experimental data for the UAH process are incomplete.
- The experimental performance of the UAH reverse osmosis unit does not support the performance assumption made in the process design.

- Economic analyses prepared by both developers appear to rely on overly optimistic estimates for capital costs and revenues from byproducts and recycling. Increased paper recycling efforts will reduce the quantity of hydrolyzable cellulose in the waste stream, and increased glass, plastic, and metal recycling efforts will reduce the byproduct credits. Moreover, increasing landfill tipping fees favorably affect the economics of MSW processing.
- The processes need to be demonstrated using an MSW feedstock representative of current, actual conditions.

At their present states of development, neither the TVA nor UAH process is ready for expansion to an economically viable commercial-scale application. Both processes are advanced enough, however, to serve as a combined base for detailed engineering and technology development studies leading to design, construction, and operation of a commercial demonstration plant, specifically:

- Combustion tests should be conducted with the residue generated from enzymatic as well as acid hydrolysis to evaluate the level of acid gas emissions that will occur.
- Detailed environmental impact assessments should be prepared before developing any further larger-scale process, to identify areas of environmental sensitivity.

## Economic Evaluation

These plants are essentially waste-processing projects, the primary objective of which is to reduce the waste stream that must ultimately be landfilled. Their development should be approached from this perspective. They should not be considered to be strictly ethanol- or energy-producing plants, even though such plants can produce both liquid and solid fuels, and electricity as byproducts.

IGT conducted preliminary economic evaluations of 400 ton/day MSW-processing plants based on either the dilute acid or the enzymatic hydrolysis processes. An important difference exists between this analysis and those performed by TVA and UAH. This analysis assumes U.S. Government ownership of the facility, and does not consider such factors as depreciation, taxes, and insurance. This analysis also assumes that the facilities operate for a 20-year lifetime at a rate of 24 hours/day, 330 days/year. It also uses a discount rate of 4.6 percent, which is consistent with U.S. Department of Energy/Department of Defense (DOE/DOD) guidelines for analysis of federal energy management programs. The results of the economic evaluations are presented in terms of simple payback period, net present value, and internal rate of return as follows:

1. The total capital investment for the base case dilute acid hydrolysis process was \$51.1 million; the base case enzymatic hydrolysis process total capital investment was \$63.2 million.
2. The operating costs were determined to be \$8.0 million and \$7.3 million/year, respectively.
3. The tipping fee for the base case was set at \$45/ton (the U.S. average).
4. The simple payback periods for the dilute acid hydrolysis and enzymatic hydrolysis processes were 12.15 and 13.20 years, respectively.
5. The net present values were, respectively, positive \$3.1 million and negative \$1.5 million.
6. The internal rates of return were 5.3 percent and 4.3 percent, respectively.

Based on this analysis, the TVA process is marginally preferable. The anticipation of escalating tipping fees would be the primary economic justification for constructing these facilities.

## Conclusions

In general, about 75 percent of the revenue for these plants is derived from tipping fees and sale of recyclable materials. Ethanol production accounts for only 18 to 19 percent of total facility revenues. In terms of waste reduction/minimization, the UAH facility is somewhat more efficient than the TVA facility, with a net reduction of 62 percent compared with 43 percent. As a result of the heavy reliance on tipping fees for income, both processes display a very high sensitivity to tipping fees and escalation of future tipping fees.

The processes differ in that the TVA process has much larger landfill requirements, while the UAH process uses much more electricity. Neither facility benefits from the sale of electricity as originally projected, although both do offset a portion of electric requirements by self-generation with residual solid fuel.

Important to the economics of MSW to fuel ethanol plants was the long-term stability of the MSW supply and the consistency of the MSW composition fed to the process. It is likely that increasing recycling efforts by local communities will reduce the quantity of recoverable and recyclable materials in the MSW stream, which will negatively impact the required tipping fees. Similarly, the quantity of cellulosic materials in the MSW stream will also likely decline, requiring that MSW be collected from a larger radius—at increasing transportation costs—to fill the existing plant capacity. Other factors, such as the suitability of the residual fraction of MSW remaining after processing for direct combustion for power generation (e.g., the ash fouling characteristics), and the environmental impact (HCl and CO<sub>2</sub> emissions), of MSW processing plants with fuel ethanol production capability, must be addressed through the application of existing technologies.

## Recommendations

Based on these observations, it is recommended that integrated, continuous pilot plant scale testing of the processes be investigated in adequate detail at an equivalent, unprocessed MSW feed rate to the plant, of about 20 ton/day. This would include MSW classification (to produce a waste-derived feedstock) followed by pretreatment (prehydrolysis), hydrolysis (acid and enzymatic), acid neutralization/enzyme recovery, glucose/xylose fermentation, and distillation. Distillation of the fermentation beer will generate stillage bottoms that should be analyzed for suitability as animal feedstock or other use. This investigation, which could be conducted during a focused 2-year period, will ensure the logical and economic progression of both processes to commercialization.



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## FOREWORD

This report was prepared for Headquarters, U.S Army Corps of Engineers (HQUSACE), under Project 4A162721A896, "Base Facility Environmental Quality"; Work Unit T01, "Investigation of Municipal Solid Waste and Paper to Alcohol Study." The USACERL Project Manager was Dr. Mike Lin.

The research was performed by the Energy Systems Division (ES) of USACERL. The USACERL principal investigators were Dr. Mike Lin and Mr. Matthew Snyder. Part of the work was performed under contract to the Institute of Gas Technology (IGT, Chicago). Mr. Christopher Blazek was the IGT Project Manager. Dr. David Joncich is Chief, USACERL-ES. The USACERL technical editor was Mr. William J. Wolfe, Information Management Office.

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# INVESTIGATION OF MUNICIPAL SOLID WASTE TO ALCOHOL CONVERSION FOR ARMY USE

## 1 INTRODUCTION

### Background

Rising costs of nonhazardous solid waste (SW) disposal, along with strict environmental regulations, limited landfill capacity, and the difficulty in siting new incinerators and landfills have all combined to make SW disposal a costly, complex problem for Army installations. Converting SW to alcohol is one proposed way to lessen the SW disposal burden. Headquarters, U.S. Army Corps of Engineers (HQUSACE) received two proposals requesting funding for the construction of processing plants for the conversion of waste paper into ethanol, a process that could potentially reduce the SW disposal burden on Army installations. One project involved a large scale-up of a test unit developed by the Tennessee Valley Authority (TVA, Muscle Shoals, AL) that used an acid hydrolysis process, and the other involved a small plant to be built at Redstone Arsenal, AL using an enzymatic hydrolysis process. Before committing to either project, the U.S. Army Construction Engineering Research Laboratory (USACERL) was asked to evaluate the technical and economic feasibility of the two proposed processes for converting cellulosic materials—such as municipal solid waste (MSW) and waste paper—to fuel ethanol and other byproducts.

The dilute acid hydrolysis process is being developed by the Tennessee Valley Authority and is the basis for a preliminary feasibility study prepared by TVA for a site-specific location in southern Indiana. The proposed plant based on the TVA process would process 400 or 1200 tons/day\* of MSW. The recyclable fraction of the MSW would be separated in a classification step to yield a waste-derived feedstock (WDF), which would be fed to the process.

The enzymatic hydrolysis process is being developed by the University of Alabama, Huntsville (UAH) and is the basis of a feasibility study prepared by UAH and Leo A. Daly, Inc. The plant described in the UAH study would convert 16 to 27 ton/day of waste paper generated at the Redstone Arsenal in Alabama to ethanol. MSW generated at the arsenal would first be processed in a classification step to produce the waste paper stream; however, the classification step was not integral to the UAH study.

Both processes convert the sugars generated from cellulose hydrolysis to ethanol via conventional batch yeast (*Saccharomyces cerevisiae*) fermentation. The fermented beer is then upgraded to 190 proof ethanol by steam distillation. In the TVA process, the ethanol stream is processed through an anhydrous column to yield 199 proof ethanol, which is then denatured. In the UAH process, the ethanol stream is subjected to pervaporation, which selectively removes water from the azeotrope using a semipermeable membrane to yield 200 proof alcohol.

Each process responds to a specific need to minimize the amount of waste material that must be landfilled or otherwise disposed of. Each process will also help to reduce the nation's dependence on imported oil through the production of fuel ethanol as a byproduct. Moreover, it is believed that ethanol will continue to increase in value and market share as a fuel, fuel extender, or octane enhancer.

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\* A metric conversion table is provided on p 109.

## Objectives

The objectives of this study were to compare and evaluate two proposed hydrolytic processes used to convert cellulosic materials (specifically, MSW and waste paper) to fuel ethanol and other byproducts, and to independently assess the processes for technical and economic feasibility.

## Approach

A literature search was performed in the areas of the dilute acid hydrolysis process, and the enzymatic hydrolysis process, focusing primarily on two feasibility studies: the Tennessee Valley Authority's *Preliminary Feasibility Study of the Recovery and Production of Recyclables, Chemicals, and Electricity From Municipal Solid Waste for Southern Indiana* (June 1990),<sup>1</sup> and a study prepared by Leo A. Daly, Inc. and the University of Alabama, Huntsville entitled *Ethanol Production Facility: Analysis/Feasibility Study* (September 1990).<sup>2</sup> Site visits were made to the TVA and UAH laboratories, and to the UAH pilot plant (Appendix A). Flow diagrams of the two processes were constructed to clarify the comparison of the process steps. Material balances were derived for both processes, to compare the processes on the basis of same-scale operations. For both processes, the following parameters were analyzed and compared:

1. The biological processes
2. The energy balances (requirements and gains)
3. The economic assumptions costs, credits, and economic bases).

## Mode of Technology Transfer

It is anticipated that the information in this report will be incorporated into an Engineering Technical Note (ETN).

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<sup>1</sup> J.D. Broder, M.M. Bulls, R.O. Lambert, and J. W. Barrier, *Preliminary Feasibility Study of the Recovery and Production of Recyclables, Chemicals, and Electricity From Municipal Solid Waste for Southern Indiana*, Report prepared by TVA and Submitted to Lincoln Hills Resource Conservation and Development Area, Inc. (June 1990).

<sup>2</sup> M.L. Hanish, (Leo A. Daly Project Manager), *Ethanol Production Facility: Analysis/Feasibility Study*, Report prepared by Leo A. Daly and University of Alabama in Huntsville for the Redstone Arsenal (October 1990).

## 2 FEASIBILITY STUDIES REVIEW

The following sections describe the review performed to evaluate and assess the feasibility studies prepared by TVA and UAH. Each review includes a brief process description, a check of the overall process material balance as presented, an evaluation of the individual processing steps, a check of the process energy requirements, and an examination of the parameters used in the preliminary economic evaluations.

### Process Descriptions

The primary sources of information for the following process descriptions were the feasibility studies prepared by TVA and UAH, supplemented when appropriate, with information obtained from the literature review.

#### *Dilute Acid Hydrolysis (Tennessee Valley Authority)*

In the preliminary feasibility study prepared by the Tennessee Valley Authority (TVA), MSW-processing plant sizes of 400 or 1200 ton/day were considered at a site-specific location in southern Indiana. (The 400 ton/day case is described here.) A simplified flow diagram of TVA's one-stage dilute acid hydrolysis process, including front-end MSW classification, from the preliminary feasibility study is presented in Figure 1. The MSW is first classified into a recyclable/recoverable materials fraction (ferrous metals, aluminum, glass, plastics, etc.) and a cellulosic fraction. Of the 400 ton/day of MSW feed, 39.7 tons (9.9 percent) of ferrous and nonferrous metals and aluminum, 37.0 tons (9.2 percent) of glass, and 28.4 tons (7.1 percent) of plastics are recovered for recycling. Some 32.4 tons (8.1 percent) are sent to the landfill. The waste derived feedstock (WDF) fraction, 262.56 tons (or 65.6 percent), is slurried with recycled water containing sulfuric acid. Steam is added to increase the temperature to about 160 °C (320 °F). In the hydrolyzer, the acid concentration is about 2 weight (wt) percent. During the 15- to 25-minute hydrolysis period, about 38 percent of the cellulose is converted to fermentable sugars.

Lignin and unhydrolyzed cellulose are removed from the hydrolyzate by filtration. The solids are washed and filtered again, then pressed to about 50 wt percent moisture. After further air drying, the solids are burned in the boiler to produce steam and electricity. Furfural and acetic acid produced from hemicellulose hydrolysis are recovered from the slurry by a stripping column. The hydrolyzate is neutralized with lime to a pH of 5.5, which is appropriate for fermentation. Gypsum (hydrated calcium sulfate) precipitated during neutralization is removed from the hydrolyzate by filtration, then washed and disposed of at a municipal landfill. (According to TVA, there are potential markets for gypsum). If the gypsum were found to exhibit any RCRA characteristics for hazardous waste (i.e., to contain leachable metals), disposal costs could range up to \$300/ton plus transportation. The hydrolyzate is then inoculated with yeast and allowed to ferment at a temperature of about 30 °C (86 °F). Carbon dioxide produced during fermentation is collected and sold as a byproduct.

After fermentation is complete (from 48 to 72 hours or more), the ethanol-rich stream is distilled to produce 190-proof ethanol and stillage. The alcohol stream is then processed through a dehydration column (molecular sieve) to 199 proof alcohol, and denatured by the addition of gasoline. The overall yield of ethanol is 30 gal/ton of dry WDF (17.2 gal/ton of dry MSW). The stillage, which contains from 96 to 99 percent water, is dried to about 40 percent moisture in a multiple-effect evaporation process. The dried product is then sent to the boiler. TVA has estimated capital and operating costs (plus byproduct credits) for plant sizes of 400 or 1200 ton/day of MSW, which is detailed in Chapter 5 (p 63).

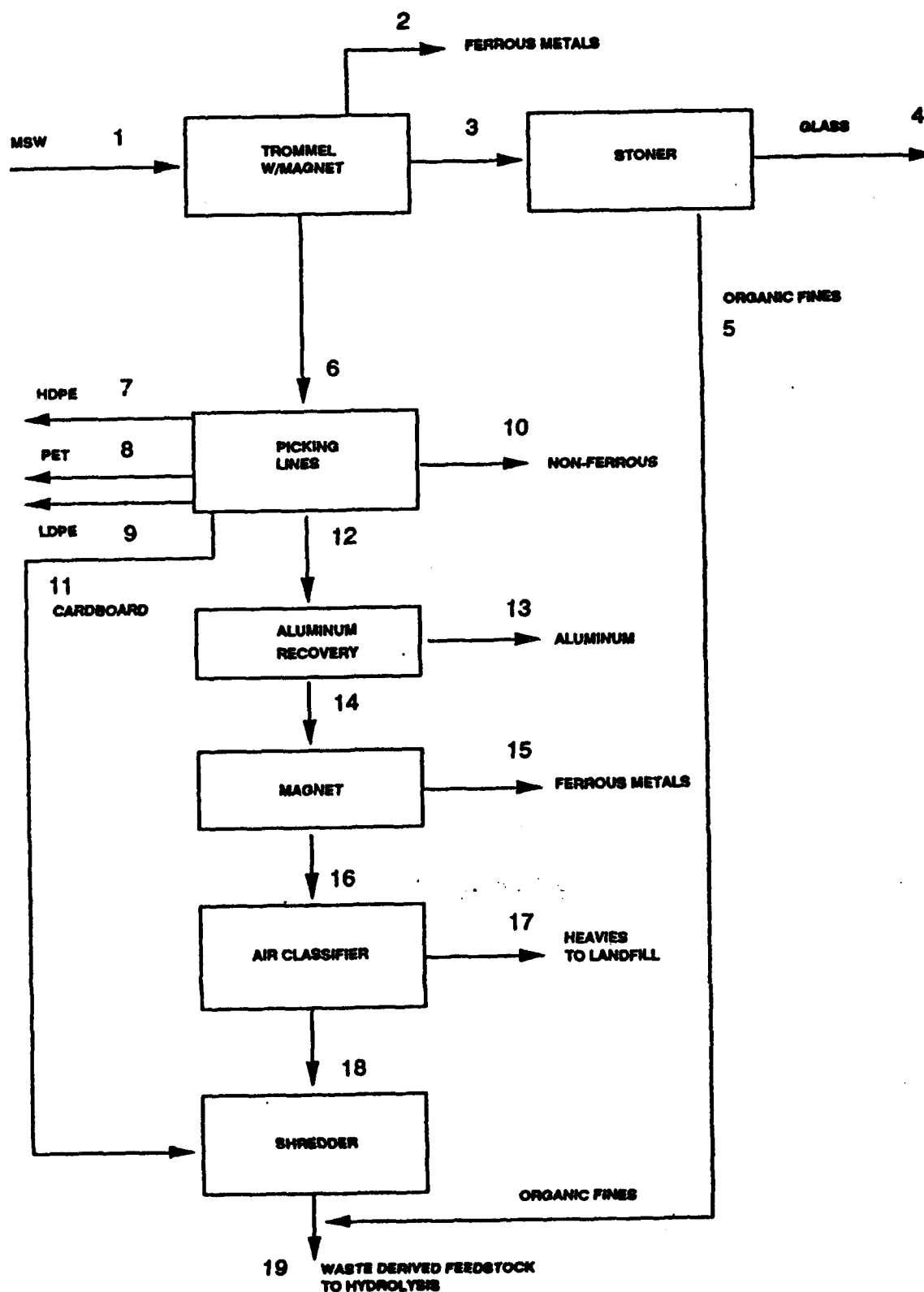


Figure 1. TVA's Single-Stage Dilute Acid Hydrolysis Process for MSW Processing.

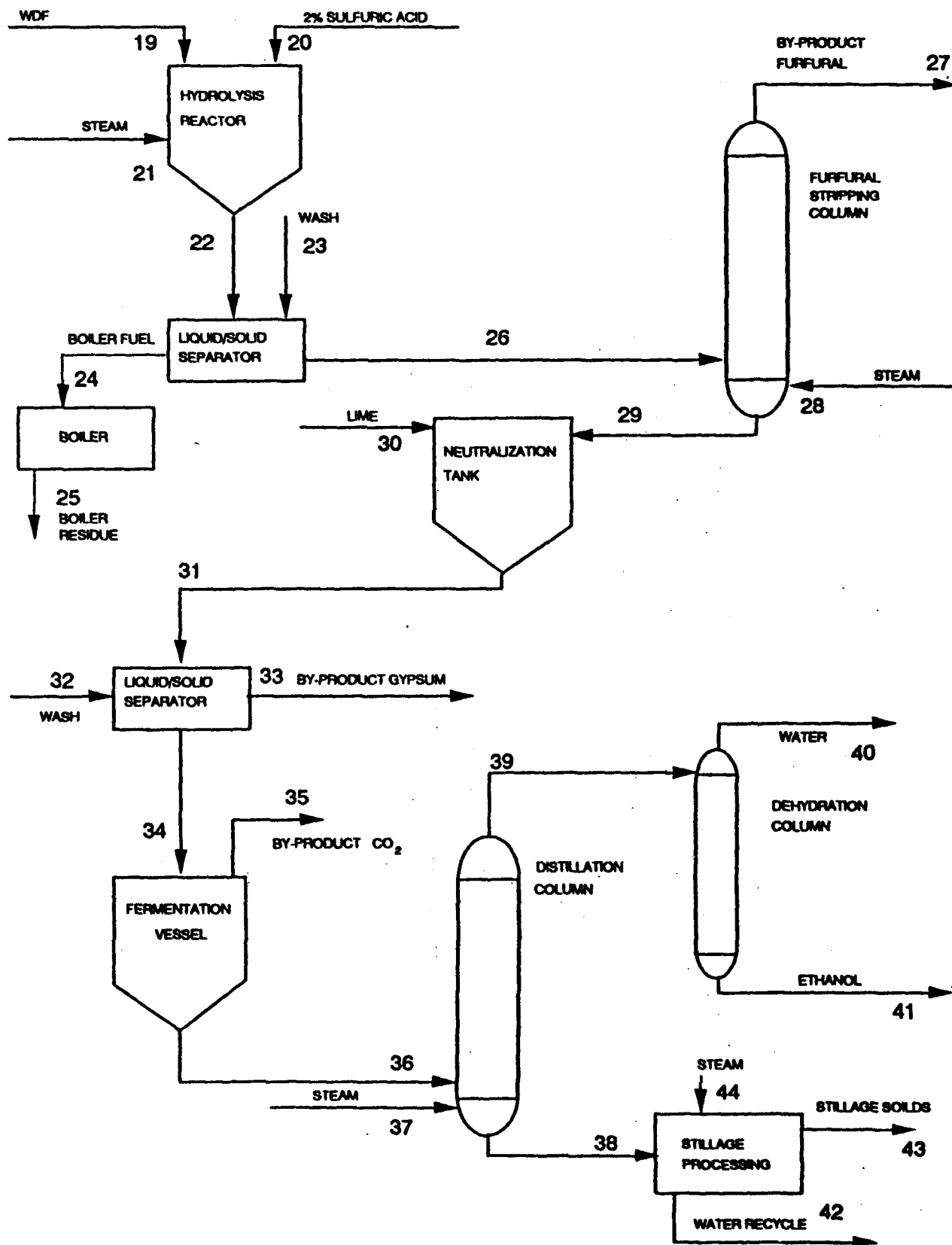


Figure 1. (Cont'd).

## Enzymatic Hydrolysis (University of Alabama in Huntsville)

In the process being developed by the University of Alabama in Huntsville (UAH), paper and paper waste will be converted to fermentable sugars by enzymatic hydrolysis. The hydrolyzate will then be fermented to ethanol with baker's yeast (*S. cerevisiae*). The proposed process is to be installed in an existing building at the Redstone Arsenal. Two scales of operation are considered: 16 and 27 ton/day. (The 27 ton/day case is described here). A simplified flow diagram of UAH's enzymatic hydrolysis process from the feasibility study is presented in Figure 2.

Shredded waste paper is loaded into the batch steam processor (by the Holloway Process) and blended with hot water (0.9 lb/lb of paper) recycled from the process. Saturated steam (50 psig) is then introduced to the reactor at 0.95 lb/lb of paper to bring the mixture temperature to  $\approx 145^{\circ}\text{C}$  ( $\approx 292^{\circ}\text{F}$ ). The mixture is held at these conditions for about 1 hour, after which the steam is vented from the processor. Some of the vented steam is used to preheat incoming hot water; the balance is conveyed to hot water storage until needed.

The sterilized paper slurry is discharged from the processor and passed over a vibratory screen to remove any particles greater than 1/2 in. The material is then charged to slurry preparation tanks and blended with recycled process water (containing the enzyme cellulase). Makeup enzyme is also added at this time. The slurry is then pumped to the hydrolysis tanks where 5 additional tons of hammer-milled paper (plus recycled water and enzyme) are blended with the original stream. In the hydrolysis tanks, the concentration of solids (paper) is about 2.85 wt percent.

Hydrolysis of the cellulose takes place at about  $55^{\circ}\text{C}$  ( $131^{\circ}\text{F}$ ) and requires 24 to 36 hours to complete. The slurry, containing glucose and other fermentable sugars, is subjected first to centrifugation and then to pressure filtration to remove the bulk of unreacted solids. The solids from both units are then washed with hot water from the steam processor (plus fresh water makeup). The solids are dried in a steam dryer and then pelletized to byproduct solid fuel with a calorific value of up to 7200 Btu/lb.<sup>3</sup> Liquids from the solids wash are combined with the glucose-rich filtrate and are conveyed to the ultrafilter. Ultrafiltration removes the enzyme from the glucose-rich stream and recycles it to enzyme recycle storage. The concentration of glucose in the filtered solution is about 1 wt percent. To reduce the amount of water that must be removed by distillation later in the process, the bulk of the water is removed by reverse osmosis (RO). At the UAH laboratory, the RO was conducted at 520 psig. More than 90 percent of the water in the stream is removed by RO. The concentrated glucose in the resulting stream is about 16 wt percent.

The glucose is then converted to ethanol by conventional yeast fermentation with *S. Cerevisiae*. Carbon dioxide produced during fermentation is collected and sold as a byproduct. Yeast produced during fermentation is also recovered and sold as a byproduct. The alcohol stream is concentrated by conventional distillation to 190 proof and then dehydrated to 200 proof using pervaporation. Pervaporation uses a semipermeable membrane to selectively remove water from the solution. The water is removed from the system in the gas phase by passing air or other gas across the membrane. Fusel oil (a mixture of  $\text{C}_3$  to  $\text{C}_5$  alcohols, mostly isopentyl alcohol) is produced by the enzymatic transformation of amino acids present in yeast. It is recovered as a byproduct of distillation and is sold as a byproduct. (According to Dr. M. Eley [UAH], very little fusel oil is expected to be produced, because waste paper does not contain any significant amounts of proteinaceous material.)

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<sup>3</sup> M.H. Eley and C.C. Holloway, "Treatment of Municipal Solid Wastes by Steam Classification for Recycling and Biomass Utilization," *Applied Biochem. and Biotech* (1988), Vol 18, pp 125-135.

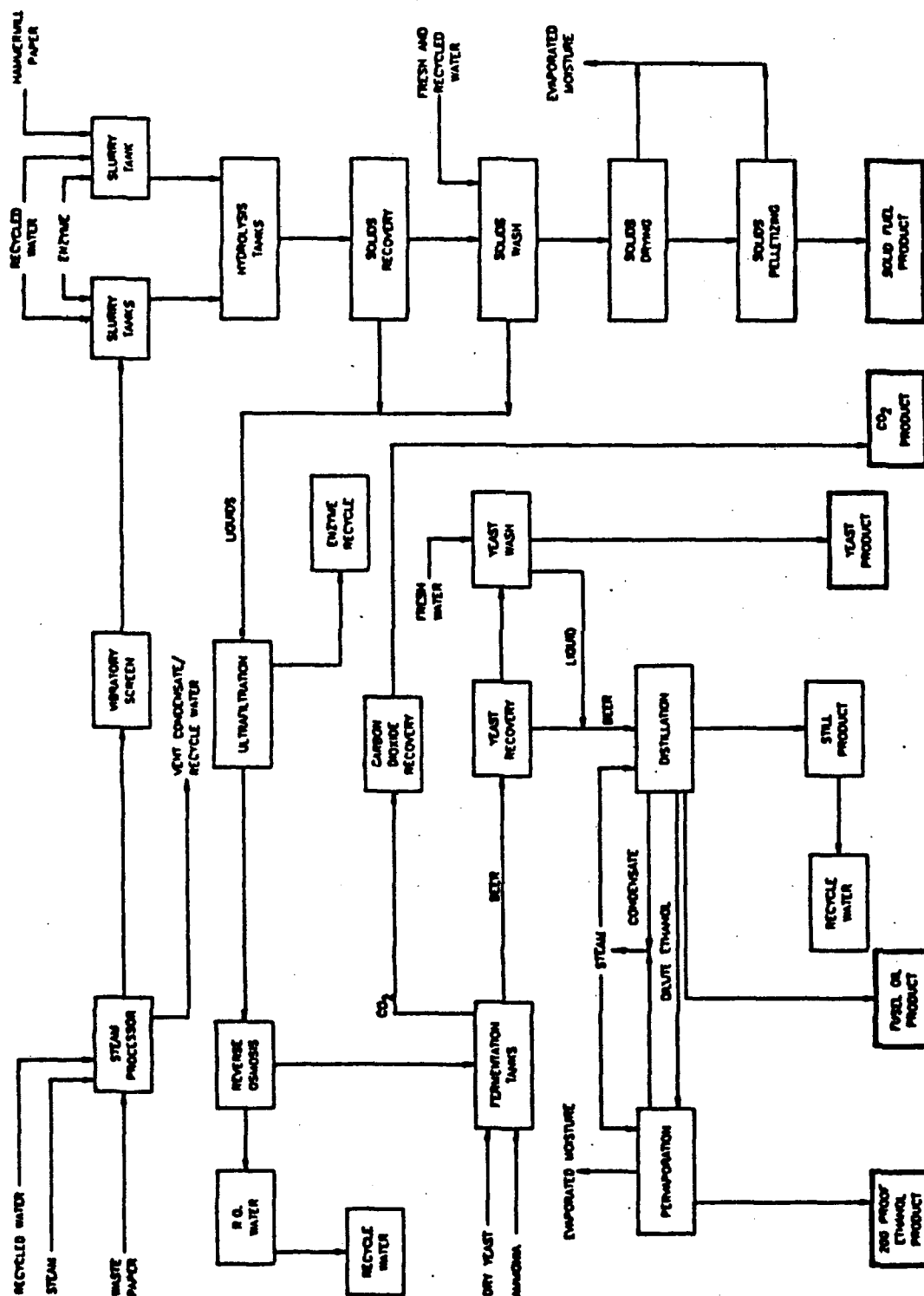


Figure 2. UAH's Enzymatic Hydrolysis Process for MSW Processing.



## Material Balance Checks

The following comments concerning process material balances are based on reviews of the feasibility studies prepared by the developers. The reviews were conducted to assure a fair comparison of the two processes based same-size operations. Questions that arose during the review concerning material and energy balances were submitted to the developers. Copies of the questions posed by and the responses submitted by the developers are included in Appendix C.

### *Tennessee Valley Authority*

The experiments to determine the operating conditions for sulfuric acid hydrolysis of RDF were conducted at TVA in a 20-liter rotating digester at temperatures ranging from 130 to 180 °C (265 to 356 °F). The results showed that the ethanol yield from newsprint could be as high as 23 gal/ton of dry feedstock. Two types of RDF tested by TVA yielded 6 and 14 gal of ethanol/ton. The ethanol yield reported in the material balance for the 400 ton/day plant was 114.3 lb or 13.6 gal/ton of dry MSW (199.5 lb or 30 gal/ton of dry WDF). Selection of this ethanol yield is considered somewhat optimistic, but justifiable for a preliminary feasibility study. TVA indicated that the quality of the RDF used for the experiments was poor. It should be noted that increased recycling activities in local communities will likely reduce the flow of paper and cellulosic material in the MSW stream, possibly reducing the quality of the MSW for ethanol production. TVA has converted a 2 ton/day system from wood waste hydrolysis to RDF and waste paper hydrolysis. Digesters for converting hydrolyzate to ethanol are being installed<sup>4</sup> but are not connected to the hydrolyzer.

Specific comments to the TVA feasibility study are as follows:

1. The concentration of sulfuric acid required for hydrolysis as described in the TVA study (p 19) is 2 wt percent; that presented in either the 400 or 1200 ton/day material balances is 3.24 wt percent (Stream 20), which is diluted to 1.5 wt percent in the total solution (2.0 wt percent if based on water and acid alone).

2. The quantities of water used for washing the residue from the Hydrolysis Reactor (Stream 23) and the gypsum byproduct from the Neutralization Tank (Stream 32) are 1539 and 638 lb/ton of dry WDF, respectively. The values are specified in the text (p 25 for Stream 23; p 23 for Stream 32). The bases for these ratios are not provided.

3. Based on the information in Tables 16 and 17, the net production of ethanol (Stream 41) is calculated to be 89.3 lb/ton of as received MSW processed, or 199.5 lb/ton of dry WDF, or 30.3 gal/ton of dry WDF. This production rate was specified by TVA for the base-case analysis (p 23) and corresponds to a net conversion of 390 lb of sugar (glucose)/ton of dry WDF. This value was also selected by TVA for the base case analysis (p 20) and is based on laboratory hydrolysis tests in which total sugar yields ranging from 160 to 390 lb/ton of dry RDF or newsprint were achieved (p 20). However, the gross amount of sugar produced from cellulose and hemicellulose hydrolysis (Stream 22) reflects a yield of 476 lb/ton of dry WDF. This includes 53.4 lb of xylose, which is fermented to ethanol by *Tannophilus pachysolen*. About 11.4 percent of the net ethanol produced is derived from xylose fermentation. The difference between the gross and net yields is accounted for in processing losses.

4. On p 29, the steam requirements for the 400 ton/day plant are mentioned as  $4.0 \times 10^{-7}$  Btu/hr; this should read  $4.0 \times 10^7$  Btu/hr. This typo is continued in Table 14 (p 30).

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<sup>4</sup> Phone conversation between Mr. Kenneth Griggs (USACERL) and Mr. Donald Walter (DOE). Subject: Proposed Waste to Ethanol Demonstration Facility (7 March 1991).

5. The energy given up by condensing the steam ( $4.02 \times 10^7$  Btu + 34,202 lb = 1175.4 Btu/lb) appears to be high unless superheated steam is used. For steam (nominal 160 psia) to provide this much energy, it must be superheated to a temperature of 965 °F. It is more reasonable to assume that the steam requirement has been underestimated. To provide  $4.02 \times 10^7$  Btu/h, a total of 46,800 lb of 160 psia-saturated steam must be condensed, or 37 percent more than that specified by TVA.

6. The material balance on p 35 shows that 37.70 tons of glucose are produced from hydrolysis of 37.7 tons of cellulose (Streams 19 and 22). The stoichiometry of the hydrolysis reaction indicates that per lb of cellulose hydrolyzed, 1.11 lb of glucose are formed. Thus, 37.70 tons of cellulose would yield 41.89 tons of glucose upon hydrolysis, or conversely, if 37.7 tons of glucose were produced, then 33.93 tons of cellulose would have been hydrolyzed.

7. On p 76, the range of laboratory sugar yields is stated as 160 to 580 lb/ton of dry RDF. The range should be more closely defined. On p 78, TVA recommends that large scale tests be conducted to determine processing conditions and product yields within  $\pm 10$  percent.

8. The process flow diagram and material balances do not include input streams for yeast and nutrients, or an output stream for byproduct yeast produced.

9. For the 400 ton/day MSW case, the solid byproduct used in the boiler to generate electricity would raise  $\approx 60,000$  lb/h of steam. This may be too small for economic electricity production.

10. Some material balances are violated. A copy of the process flow diagram and the corrected material balance tables are included in Appendix B.

11. There are species missing from several intermediate streams. However, all species are confirmed in the WDF (Stream 19) except water. A balance around the "Shredder" shows that the amount of water in Stream 19 should be 48.43 tons (yielding a moisture content of 21.8 wt percent) instead of 83.48 tons (yielding a moisture content of 31.8 wt percent). This amount of water in Stream 19 has apparently been adjusted to conform to the moisture content of the WDF composition given in Table 5 (p 18).

12. The bases of steam feeds to the furfural stripping column (No. 28), distillation column (No. 37), and stillage processing (No. 44) are not known. These steam flows do not show up in product streams 27 and 29, 38 and 39, or 42 and 43, respectively. Since the purpose of this steam is to provide heat only, the quantity of condensate leaving the column should be shown to reflect material balance. This was confirmed by TVA (Appendix C).

13. The material balance table shows that about 68 percent of the hemicellulose entering the "Liquid/Solid Separator" stage via Stream 22 is converted to acetic acid. It would be important to determine whether acetic acid formed in this separator or in the "Hydrolysis Reactor."

14. On p 26, in the paragraph beginning with "Gypsum," the RCRA corrosivity analyses for solids requires the pH to be *less than* 2.0 (not above) and *greater than* 12.5 (not below).

15. An air feed to the "Boiler" and flue gases leaving the "Boiler" should be shown.

16. Some of the sulfuric acid remains with the dewatered solids that will be combusted in the boiler. There could be a problem with acid gas condensation in the stack unless the acid is adequately neutralized. Similarly, the levels of hydrochloric acid (chlorine is present in some plastics) and sulfur dioxide (SO<sub>2</sub>) in the stack must be controlled to conform to regulations.

## *University of Alabama in Huntsville*

The process being developed at the University of Alabama in Huntsville (UAH) uses an enzyme (cellulase) to hydrolyze the cellulose in the waste paper to fermentable sugars, which are then biologically fermented to ethanol. Other byproducts are carbon dioxide, fusel oil, yeast, and a solid fuel. The enzyme is recovered and recycled. The Analysis/Feasibility Study prepared by Leo A. Daly for UAH omitted material that UAH considered proprietary. Two references were found in the literature that described either the patented Holloway process for steam sterilizing the MSW before it is classified, or the UAH enzymatic hydrolysis process. Overall, very little experimental data were available to evaluate the UAH process.

The scales of operation considered in this study by UAH for producing ethanol from waste paper are 16 and 27 ton/day. The plant would be sited in a specific building (No. 5410) at the Redstone Arsenal. UAH could then possibly enter the newsprint/paper recycling market with their process.

Specific comments to the UAH feasibility study are as follows:

1. The net ethanol yield reported in the UAH material balance for the 16 and 27 ton/day plants was 390.3 lb/ton of waste paper.

2. The material balance tables provide "IN" streams for several unit operations without any "OUT" streams. At steady state, the amount going into a subsystem should be equal to that going out of the subsystem and should be so designated.

3. No experimental data are provided to support conversions. No descriptions of experimental equipment (scales of operation tested) are provided. For example, what are the bases for the design of the hydrolysis reactors and the fermentation vessels?

4. Each process stream should be given a unique stream number for proper identification. Streams numbers 3, 4, 5, 6, 20, 25, 37, and 45 have been used more than once.

5. Several important streams, including those leaving the system, have not been numbered (such as fusel oil recovered from the Distillation step, Steam entering the Fuel Dryer, condensate leaving the Fuel Dryer, etc.).

6. By checking the material balance around each box (when possible), material balance violations have been identified at several places. A copy of the process flow diagram and the corrected material balance tables are included in Appendix B.

7. To balance, the fusel oil make for the 16 ton/day case should be 557 lb instead of 443 lb. Thus, the fusel oil produced is 557 and 573 lb for the 16 and 27 ton/day cases, respectively. These values are not in the expected proportion of 27/16 or 1.6875. Furthermore, the reported amount of fusel oil produced appears to be high. This observation was later confirmed by Professor M. Eley (UAH), who agreed that little fusel oil is expected by fermentation of the paper waste because it contains little or no proteinaceous material.

8. The "net daily material balance" table has errors and the steam and water feeds shown are incorrect. All recycle streams are internal streams and do not leave the system. Recycle streams should not be included as "OUT" streams. There are two condensate streams leaving the system, which must be included. Fusel oil make is not included.

9. In the "Solid Fuel Recovery" balance table, the steam entering the subsystem and condensate leaving the subsystem are not shown. In the "Ethanol Recovery" balance table, the amount of "Still Bottoms" leaving the subsystem is shown incorrectly. Also, the fusel oil leaving the ethanol recovery subsystem is not shown.

10. The enzyme feed rates to the upper slurry tank are different (176,339 and 190,958 lb), although the hammer-milled paper feed rate is the same for the two cases. These should be identical. The enzyme feed to the lower slurry tank should be in the ratio of 1:2; the same as that of the feeds to the steam processor.

11. The system requires 46,895 and 67,795 lb/hour of steam for the two cases, respectively. The study does not indicate how steam is to be generated.

12. In the fuel dryer, 1 lb of steam is added to evaporate 1 lb of water. There should be some allowance for heat losses and inefficiency.

13. The compositions of the two types of feed are not specified. The basis of designs for various stages are not provided, such as residence times, concentrations, temperatures, pressures, etc. Items related to these factors cannot be checked.

14. The trip notes indicate that the reverse osmosis step concentrates the 1 to 2 percent sugar solution into 9 to 10 percent sugar solution. The UAH balance shows that the stream leaving reverse osmosis unit contains about 16 percent sugar.

15. The recycle still bottoms stream contains some dissolved solids not removed by filtration. These solids should be purged from the system at steady state.

16. The yeast byproduct represents a seven-fold increase in biomass compared to the Dry Yeast stream. In typical yeast fermentations, the yeast biomass triples.

## **Biological Processing Steps**

### ***TVA Process***

The success of the fermentation step in the TVA process depends upon the chemical composition of the hydrolyzate. Before any scale-up to a commercial size demonstration plant is considered, the following areas of concern need to be resolved in the present pilot plant.

In the laboratory scale studies, the maximum total sugar concentration that could be achieved was 3.5 percent (estimated maximum fermentable sugar concentration would be ~2.8 percent), and the maximum ethanol concentration achieved was 1.3 percent. The base-case process for the TVA feasibility study assumes a fermentable sugar concentration of about 7 percent and an ethanol concentration of about 3.5 percent in the fermenter. These higher concentrations of fermentable sugars and ethanol assumed for the base case are probably the minimum that need to be achieved for economical ethanol recovery, but apparently have not been demonstrated in the laboratory or the pilot plant. According to TVA (p 22), "When WDF hydrolyzate sugars are concentrated, the organic and inorganic components in the hydrolyzates will also be concentrated and may inhibit fermentation."

Laboratory and pilot plant studies indicate broad variation in hydrolyzate composition and sugar concentrations in different runs. Such broad variations in hydrolyzate composition and sugar concentration most likely will upset the fermentation process.

Ash and metallic components of the feed material are expected to generate soluble salts as a result of the acid hydrolysis. Inhibition of yeast activity should be expected according to the type and concentration of metallic ions, if these ions end up in the fermenter.

### *UAH Process*

The UAH process brings the sugar concentration to levels typically applied in ethanol fermentations. There does not appear to be any problem in the scale-up of the fermentation step in the UAH process.

Points of concern in the UAH process are the sensitivity of the enzyme reaction to cleanliness and purity of the paper waste; potential inhibition of the enzyme by metal ion contaminants; potential difficulties in maintaining sterile conditions in the enzyme reactor and in the reverse osmosis and ultrafiltration units.

### **Energy Requirements Checks**

The material balances presented by TVA (TVA Tables 16 and 17) do not indicate the condition of the process streams (i.e., temperature and pressure). Accurate energy balance checks cannot be made without this information. Similarly, the material balances prepared by UAH for the two scales of operation do not indicate the condition of the process streams. Again, accurate energy balance checks cannot be made without this information. Because the conditions of the process streams were not explicitly provided, the energy checks were limited and approximate. Only the major energy-intensive components, such as the UAH steam processor and the beer distillation units for both processes, were checked.

The energy balance around the UAH steam processor was determined to be incorrect as reported for the 22 tons of waste paper/day case. (Thus, adjustments in the steam and hot water feed rates would be required for scale-up.) The energy requirement check showed that the heat provided by the steam feed far exceeded that required to heat up the solids and hot water feed. Instead of 41,800 lb of steam for the 22 tons of waste paper feed, researchers' estimate indicated that 29,600 lb of steam would be sufficient.

The UAH literature indicates that in the steam processor, about one-third of the steam feed is condensed, while two-thirds is vented. The steam vented in the UAH balance represents 35 percent of the steam feed rather than 67 percent.

The required steam feed to the steam processor was estimated by energy balance. Based on the UAH literature, it was assumed that one-third of the steam feed was condensed to heat the waste paper in the processor, while two-thirds escaped as steam during depressurization. The condensed steam becomes part of the solids leaving the steam processor (Stream 4).

In the UAH process flow diagram, the total steam required for the distillation step to obtain 200 proof ethanol is 3.46 lb/lb of ethanol produced. This quantity is consistent with the literature value for advanced distillation processes—thus this value has been retained in the present balance. Pervaporation would not require the use of a third component, such as benzene or hexane, to affect the separation of the water and ethanol, and by virtue of this, may be more environmentally acceptable.

In the TVA process flow diagram, the total steam required for the distillation plus dehydration step to obtain 200 proof ethanol (from 3.29 wt percent ethanol) is 3.33 lb/lb of ethanol produced. This quantity is consistent with the literature value for advanced distillation processes—thus this value has been retained in the present balance.

### Economic Assumptions and Bases

The objective of the economic evaluations performed by the proposing developers was to demonstrate that their technologies are economically viable and yield a valuable product in fuel ethanol, while at the same time using a feedstock that would most likely be landfilled at the expense of increasing tipping fees. From the standpoint of local municipalities, waste-to-ethanol processes should be considered as waste minimization/waste reduction processes rather than energy processes. This consideration also holds for "waste to energy" (WTE) processes. Ultimately, the economics should be reckoned on this basis and the scale of operation should be appropriate for each locale. From the standpoint of military bases and reservations, local control of the waste stream composition and mandatory separation of wastes at the source may eventually render smaller facilities (such as the one proposed for the Reustone Arsenal) economically viable.

MSW classification processes have been demonstrated at different scales of operation at a number of sites across the United States. The State of Illinois has prepared a list of costs to construct MSW classification plants of sizes less than 550 ton/day. The costs range from \$200,000 for an 80 ton/day plant to \$4.06 and \$6.08 million for two 500 ton/day plants. One 450 ton/day plant was reported to cost \$30.4 million to construct in Philadelphia. The degree of classification that the MSW is subjected to at these plants may not reflect the specific process requirements of the TVA or UAH processes for WDF (waste derived feedstock).

#### *Tennessee Valley Authority*

Economic evaluations were made by TVA for MSW processing plant sizes of 400 and 1200 ton/day. For the 400 ton/day plant, the total plant investment (TPI) required was estimated to be \$35.08 million with total annual revenues of \$6.81 million. For the 1200 ton/day plant, the TPI required was estimated to be \$74.18 million with total annual revenues of \$20.44 million.

Another reference by TVA published in 1991<sup>5</sup> estimated the TPI for a 100 ton/day plant to be \$104.1 million. Furthermore, TVA projects the cost to construct a 100 ton/day waste paper/RDF (refuse derived feedstock) to ethanol pilot plant at \$53 million, or at about 50 percent less than the cost of the 1000 ton/day plant (TVA trip notes).

Note 1. The cost of the front-end classification system was estimated (per vendor quote for the entire system) to be \$1,485,148 and \$4,455,446 for the 400 and 1200 ton/day plants, respectively.

Note 2. Chemicals generated during the process as products or byproducts, include ethanol, furfural, acetic acid, and CO<sub>2</sub>. The total projected revenue from the sale of these items is \$24.42/ton. Byproduct credits assumed for the recovery of recyclable materials from the MSW received at the process plant total \$23.69/ton and represent 45.8 percent of the total revenue. The sale of ethanol and aluminum (for recycle) represent 31.4 percent and 24.4 percent, respectively, of the total revenues for both size plants. Thus,

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<sup>5</sup> M.M. Bulls, T.M. Shipley, J.W. Barrier, R.O. Lambert, and J.D. Broder, "Comparison of MSW Utilization Technologies—Ethanol Production, RDF Combustion, and Mass Burning," *Southern Biomass Conference* (Baton Rouge, LA, 1-7 January 1991).

changes in the markets for recyclable materials would significantly impact the overall economics of the plant.

*University of Alabama in Huntsville*

Capital costs for installing the equipment in existing facilities at the Redstone Arsenal were estimated to be \$4.12 million and \$5.94 million for 16 and 27 ton/day plant sizes, respectively.

The economic evaluation of the UAH process was based on the use of an existing building (No. 5410) at Redstone Arsenal. Minor repairs and modifications are required to the building to render it suitable for the intended application. Leo A. Daly considered four alternative scenarios in the evaluation including: (1) scale of operation (16 and 27 ton/day of waste paper feed); (2) use of Building 5410 or construction of a new one; (3) installation of a boiler; or (4) not installing a boiler. The evaluation determined that the use of existing Building 5410, with a 27 ton/day of waste paper feed, and no boiler would be the most cost effective.

During the 24 April 1991 trip to the UAH facility, researchers received the information that the Redstone Arsenal would provide the front-end classification system required to separate waste paper from the overall solid waste stream. For the UAH/Leo A. Daly study, it was assumed that process steam requirements would also be met by the Redstone Arsenal steam plant at no cost.

Note 1. There is an error on the summary table showing the Capital Equipment Cost Estimate for the 16 ton/day case from the UAH/Leo A. Daly study. The auxiliary costs subtotal shows \$12,000,000 (typo for \$1,200,000) and should read \$1,500,000. Correcting the error and following through, the total estimated capital equipment plus auxiliary costs should be \$4,042,800 with a 10-percent contingency of \$404,280 to yield a total of \$4,447,080.

Note 2. There is also an error on the same table for the 27 ton/day case. The auxiliary costs subtotal shows \$1,360,000 and should read \$1,270,000. Correcting the error and following through, the total estimated capital equipment plus auxiliary costs should be \$5,313,600 with a 10-percent contingency of \$531,360 to yield a total of \$5,844,960.

Note 3. Major components for the 27 ton/day plant account for 80.1 percent of the total capital equipment cost. These items include the steam processor (4.95 percent), the hydrolysis tanks (7.9 percent), the centrifuge (3.7 percent), the ultrafiltration units (20.3 percent), the reverse osmosis system (17.3 percent), pervaporation system (19.8 percent), and the solid fuel dryer and pelletizer (6.2 percent). Had the solid fuel boiler and turbine been included in the selected scenario, the total capital cost would have been increased by \$1,452,000 (\$1,320,000 plus \$132,000 contingency).

Note 4. It should be noted that the auxiliary costs for the 16 ton/day plant and the 27 ton/day plant contain inconsistencies. For example, the auxiliary costs for the smaller plant are greater than those of the larger plant. The cost of catwalks for the smaller plant is \$40,000; for the larger plant, catwalks cost \$10,000. The installation cost for the smaller plant is \$300,000; for the larger plant, installation is \$200,000. Engineering for the smaller plant is \$200,000; for the larger plant, engineering is \$100,000. Freight for the smaller plant is \$400,000; for the larger plant, freight is \$100,000.

Note 5. No costs for steam are included in the cost estimate (p 13). Even if the fuel is free, the cost to generate steam is not. Further, no costs for makeup enzyme are included in the cost estimate (p 13). This is necessary for the economic evaluation.

Note 6. No basis for the cost estimate of the Pervaporation Subsystem is given.

**Note 7.** Reasons for selecting pervaporation over more conventional dehydration processes are not given.

### **Summary of Feasibility Studies of Costs, Credits, and Economic Bases**

The cost of raw materials, utilities, and byproduct credits taken for the two processes are presented in Tables 1 and 2. The byproduct credits and process utility costs for the TVA and UAH economic evaluations are different and largely not comparable.

TVA calculated the required tipping fee to achieve a 15 percent return on investment (ROI). For the base-case scenario, this cost was \$57.97/ton. The UAH economic evaluation used the local tipping fee of \$28.00/ton for the waste paper feedstock for their process. Sulfuric acid costs were included by TVA, but the cost of enzyme and enzyme makeup for the UAH process was not provided.

The utilities costs for the TVA process included lime for acid neutralization as well as separate costs for process water and cooling water. UAH utility costs include purchased electricity, fuel, and chemicals. In the UAH scenario with a boiler and turbine included, the chemicals cost was \$30/ton. The cost of process water at the Redstone Arsenal is almost four times higher than that assumed for southern Indiana in the TVA process.

The ethanol prices listed by TVA and UAH are \$1.15 and \$1.55/gal, respectively. Neither developer calculated the cost to produce ethanol; both selected the going rate to calculate the required tipping fee (TVA) or the net present value of the process (UAH).

Byproduct credits are taken by TVA for electric power generated by the combustion of solids that remain after WDF hydrolysis, while UAH must sell the solid fuel as a byproduct. Credits are also taken by TVA for acetic acid, furfural, and carbon dioxide. UAH takes credits for solid fuel, yeast, fusel oil, and carbon dioxide. The only byproduct credit in common to the two processes is that for CO<sub>2</sub>, which is \$10/ton in the TVA case and \$60/ton in the UAH case (Table 2).

For the base-case scenario, TVA projects the total revenue from the sale of byproduct chemicals to be \$24.42/ton. Byproduct credits assumed for the recovery of recyclable materials from the MSW received at the process plant total \$23.69/ton and represent 45.8 percent of the total revenue. The sale of ethanol and aluminum (for recycle) represent 31.4 percent and 24.4 percent, respectively, of the total revenues for both size plants. Thus, changes in the markets for recyclable materials would significantly affect the overall economics of the plant. The UAH economic evaluation did not include any front-end processing or recovery of recyclable materials.

Table 3 summarizes the factors involved in the two developers' economic evaluations and includes stream factor and other relevant economic parameters; the economic evaluations for the two processes are not comparable. Major changes in assumptions would be required to create a common economic basis for the two approaches.



Table 1

## Raw Materials and Process Utility Costs

Type	Amount	
	TVA	UAH
<b>Raw materials costs</b>		
Enzyme, lb	NA*	NP**
MSW, ton	(57.97)	NA
Paper/waste paper, ton	NA	(28.00)
Redstone Arsenal		
Landfill Impact, ton	NA	(17.64)
Sulfuric acid, ton	70	NA
<b>Utilities</b>		
Chemicals, per ton of paper	NA	25 w/o boiler
Electricity, kWh	--	0.056
Fuel (diesel), gal	--	0.056
Lime	45	--
Steam, 1000 lb	--	0.00
Water, 10 <sup>6</sup> gal		
Cooling	50	--
Process	500	1984

\* NA = Not applicable.

\*\* NP = Not provided.

Table 2

## Byproduct Credits

Type	Amount (\$)	
	TVA	UAH
<b>Product</b>		
Ethanol, gal	1.15	1.55
<b>Byproduct</b>		
Acetic acid, ton	580	NA
Carbon dioxide, ton	10	60
Electricity, kWh	0.04	--
Furfural, ton	800	NA
Fusel oil, gal	--	6
Solid fuel, ton	--	13
Yeast, lb	--	1
<b>Recyclables</b>		
Aluminum, ton	1000	--
Ferrous metals, ton	50	--
Glass, ton	15	--
Plastics, lb	140	--

Table 3

## Economic Evaluations Factors

Factor	Cost (\$)	
	TVA	UAH
Operation	90	
Stream factor, %		100
Maintenance and Repair, %	--	
of TPI	--	10
Persormel, FTE		11.6
Economic factors		
Discount rate, %	--	7.4
Period of analysis, years	20	25
Base year	1990	1991
Inflation rate, %		
1992	0	4.0
1993		3.7
1994		3.4
1995-2015		3.1
Depreciation*	SL+	DDB++
Return on investment, %	15	--

\*+ Straight line.

++ Double declining balance.

### 3 PROCESS COMPARISON AND ECONOMIC EVALUATION ON A COMMON BASIS

One of the primary objectives of the study was to prepare revised material balances for the acid and enzymatic hydrolysis processes and to evaluate the economics of each under conditions that could be considered comparable. This meant that an appropriate plant size had to be selected to scale the two processes to a common size. The economics of each plant were then to be reevaluated and compared.

#### Scale of Operation

The developers of the dilute acid hydrolysis process (TVA) evaluated plants that processed either 400 or 1200 ton/day of MSW. The developers of the enzymatic hydrolysis process (UAH) considered waste paper processing plant sizes of 16 and 27 ton/day. It seemed logical to scale up the enzymatic hydrolysis process to include a front-end MSW classification step, rather than scale down the dilute acid hydrolysis process. In this way, the emphasis could be placed correctly on MSW processing rather than on producing ethanol for fuel.

According to Hocker,<sup>6</sup> the average waste handling capacity of all WTE plants brought on line prior to 1988 was 574 ton/day. Recently, however, the average plant size has been 1056 ton/day. This could indicate that economies of scale and return on investment warrant plant sizes in the 1000 to 1500 ton/day range. However, others believe that future WTE plants will be in the 300- to 800-ton/day range, because that size is appropriate for many local communities. Also, because of community recycling efforts, the feedstock for WTE plants will become more like that produced from plants that produce RDF from MSW.

The selection of the appropriate generic plant size devolved on either 400 or 1200 ton/day, either of which fall within the ranges mentioned in the Hocker article (300 to 800, or 1000 to 1500 ton/day). Both scales of operation had been essentially considered by TVA. Based on the Hocker article, researchers selected a base case MSW processing plant size of 400 ton/day for the generic evaluation.

#### Feedstock Characteristics

Because the generic process scale-up and economic evaluation would be based on MSW rather than waste paper, a common feedstock composition was required. As the feedstock to the UAH process was paper and waste paper (essentially 100 percent cellulosic materials), the composition of the material being fed to the ethanol processing plant was fixed to that described in the TVA feasibility study. Four hundred tons/day of MSW would be fed to the front-end classification section in each plant. The composition of the MSW (as received and dry basis) is shown in Table 4 from the TVA study (p 8) and represents the U.S. national average. The quantity of recoverable and recyclable materials separated in this step would be the same as that indicated in the TVA preliminary feasibility study. The resulting WDF would have the composition shown in Table 5 from the TVA study (p 18).

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<sup>6</sup> C. Hocker, "Waste-to-Energy Development: Who's Doing What and Why?", *Solid Waste & Power* (August 1991), pp 12-19.

Table 4

Composition of Municipal Solid Waste		
Component	Wt % as received basis	
Moisture	--*	21.86
Paper and paperboard	41.3	30.98
Glass	9.9	9.90
Aluminum	1.4	1.40
Ferrous/tin/steel	8.85	8.85
Nonferrous	0.25	0.25
Plastics**	10.1	10.10
Rubber and leather	3.3	2.81
Textiles	2.4	1.92
Wood	4.8	3.60
Food wastes	10.5	3.68
Yard wastes	5.1	2.55
Misc. inorganics	2.1	2.10
Total	100.0	78.14

\* Moisture is incorporated into the individual components here.

\*\* Total of HDPE, LDPE, PET, and other plastics.

## Generic Process Descriptions

Figure 3 shows a simplified block flow diagram for both the dilute acid (TVA) and enzymatic hydrolysis (UAH) processes. The major processing sections are: (1) MSW-processing facilities, (2) hydrolysis, (3) fermentation, (4) ethanol recovery, and (5) utilities. A description of the major equipment included in each section is included below.

The MSW classification section is identical for both the dilute acid (TVA) and enzymatic hydrolysis (UAH) processes, and consists of a trommel equipped with a magnet for separating large items. A stoner is used to separate glass from organic and inorganic material. Nonferrous metals and plastics are recovered in the hand-picking lines. An NRT aluminum recovery system is used to recover aluminum. Ferrous metals are separated from the waste stream in the magnetic separators. An air classifier is used to separate the "heavies" from the "lights" fraction, which comprises the waste derived feedstock (WDF). A low-speed shredder is used to shred the WDF before hydrolysis. In both cases, a total of 262.56 tons of WDF are fed to the hydrolysis reactors per day.

### Acid Hydrolysis (TVA)

Some minor modifications were made to the original TVA process flow diagram (refer to Figure 1) which are included in Figure 4. The hydrolysis section consists of an alloy-lined hydrolysis reactor where WDF is reacted with dilute sulfuric acid to produce fermentable sugars. The steam is added to the reactor for temperature control. The hydrolyzed materials are depressurized in the flash tanks. The solids residues are separated from the hydrolyzate in the filters, dewatered, and recovered as boiler fuel. The combined wash liquor and hydrolyzate containing sugars, furfural, and acetic and sulfuric acids is fed to stripped columns where furfural and acetic acid are recovered as byproducts by steam stripping. The

**Table 5****Composition of Waste Derived Feedstock (WDF)**

Component	Wt %	
	As Fed Basis	Dry Basis
Cellulose	37.8	55.4
Hemicellulose	3.8	5.6
Lignin	10.6	15.5
Ash	8.6	12.6
Other	7.4	10.9
Moisture	<u>31.8</u>	<u>—</u>
Total	100.0	100.0

liquid is then neutralized with lime in the neutralization tanks. The resulting gypsum is filtered from the hydrolyzate and then disposed of in a landfill.

The fermentation section consists of a series of fermenters equipped with agitators. The heat of fermentation is removed by the heat exchangers. The yeast culture is prepared in the prefermenter and added to the fermenters by metering pumps. The fermentation off-gases, largely CO<sub>2</sub>, are collected, purified, and recovered as a byproduct. The fermented broth is stored in a beer well prior to distillation and dehydration.

The ethanol recovery section consists of a conventional distillation column equipped with reboiler and condenser, where ethanol is recovered from the fermentation broth. The resulting azeotropic ethanol is dehydrated in a commercial azeotropic distillation system using benzene (or other material) as the entrainer. The solids residues are filtered from the stillage and recovered as boiler fuel. The water is recycled to the process.

The utilities section consists of a boiler for steam production and a turbogenerator for electricity generation. The solid residues recovered from the hydrolysis and ethanol recovery sections are burned in the boiler. A cooling tower, and storage vessels for the raw materials and products are also included in this section.

The net material balance for the plant is summarized in Table 6; the complete material balance is summarized in Table 7. Based on questions about the material balance given in the TVA study, TVA produced a revised balance. The balance shown in Table 7 is essentially the same as the revised balance provided by TVA except the following items. Appendix C contains the questions and the revised TVA material balance.)

In the hydrolysis reactor, water is added to cellulose to form glucose (one mole of water is consumed per mole of glucose formed, an increase in weight of about 11.1 percent). The revised material balance provided by TVA does not show the consumption of water by cellulose. Researchers considered two approaches for making the adjustment:

1. The amount of cellulose converted (37.70 ton/day) to glucose can be assumed to be correct. In this case, the glucose produced would be increased from 37.70 to 41.89 ton/day (Stream 22) and would result in an overall ethanol production of 33.70 gal/ton of dry WDF.

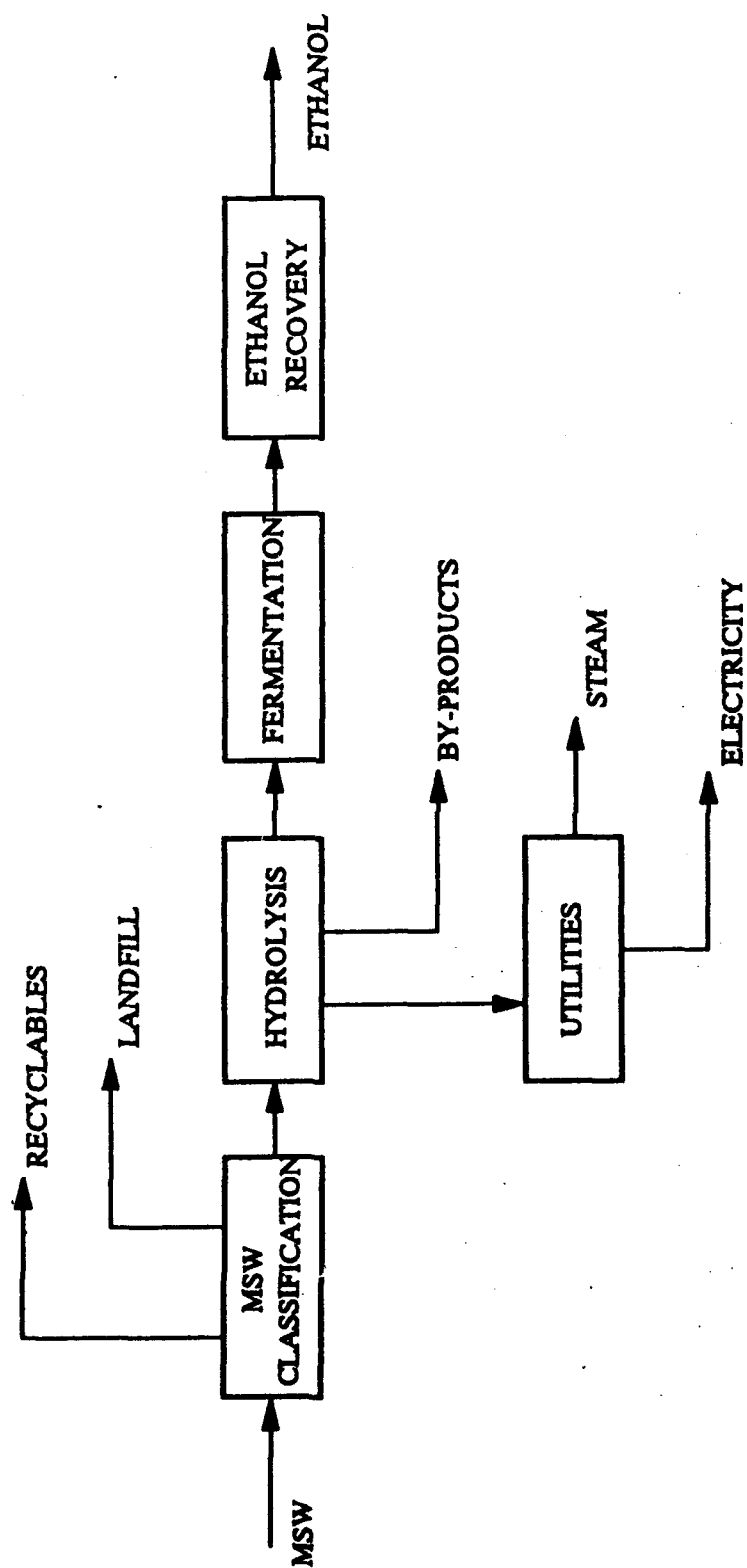


Figure 3. Flow for Generic MSW Processing Plants for Producing Ethanol.

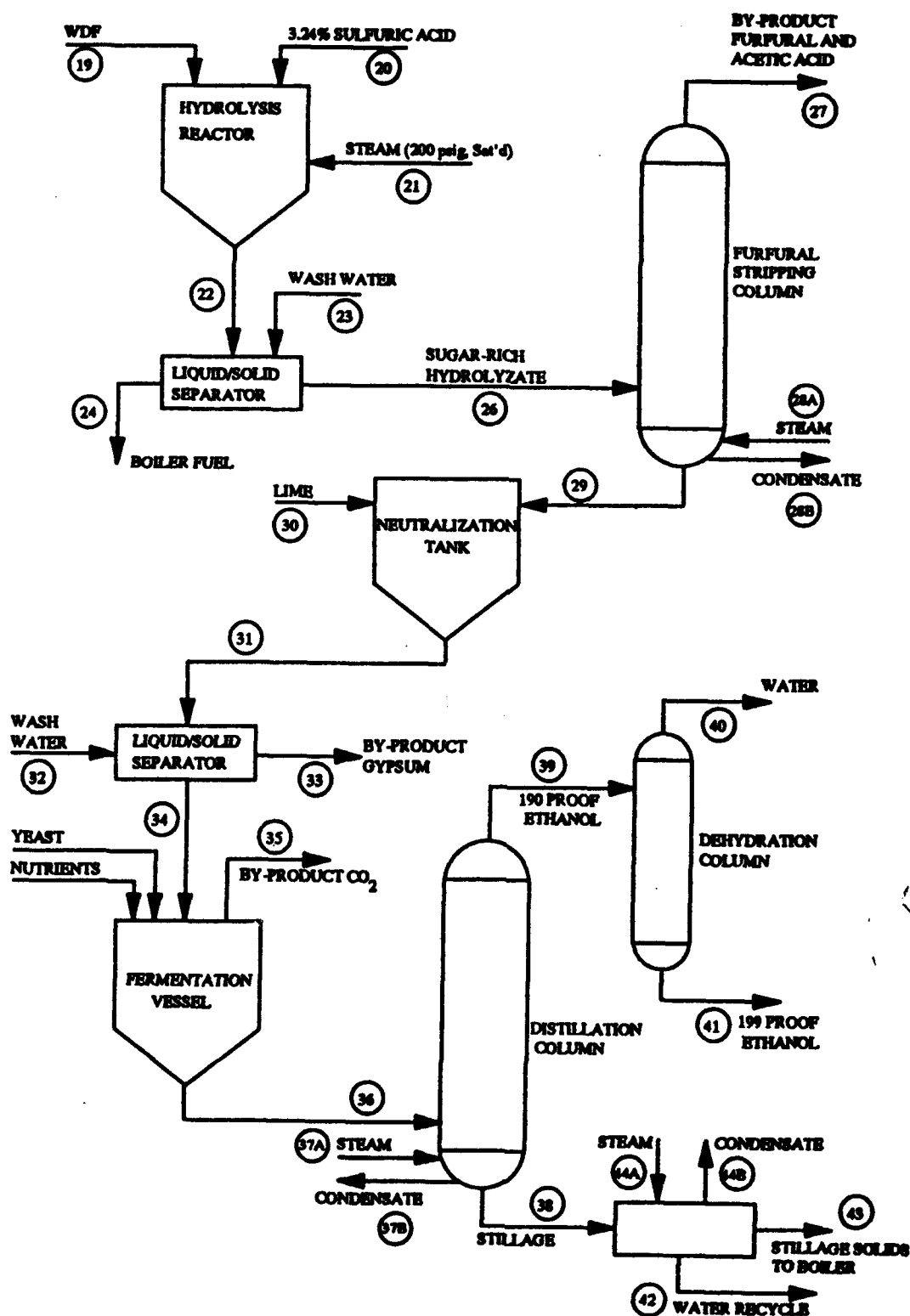


Figure 4. Process Flow for a 400 ton/day MSW (262.56 ton/day WDF) to Ethanol Plant Based on the TVA Dilute Acid Hydrolysis Process.

Table 6

Net Material Balance for Dilute Acid Hydrolysis Process  
(Based on 400 ton/day MSW or 262.56 ton/day WDF Feed)

Material	Amount
<b>Inputs</b>	
Waste-derived feedstock (WDF)	262.56
Sulfuric acid (3.24 wt %)	331.82
Steam	410.43
Wash water	193.46
Lime	<u>7.87</u>
Total	1206.14
<b>Outputs</b>	
Ethanol (200-proof) product	17.86
CO <sub>2</sub> product	17.69
Boiler fuel	266.89
Acetic acid and furfural	4.21
Condensate	288.46
Gypsum by-product (to disposal)	65.11
Stillage solids to boiler	6.68
Water	<u>539.24</u>
Total	1206.14

2. The amount of glucose formed (37.70 ton/day - Stream 22) can be assumed to be correct. In this case, the amount of cellulose converted would be reduced from 37.70 to 33.93 ton/day. This would result in an overall ethanol production of 30.32 gal/ton of dry WDF.

The TVA study states (p 23) that the basis of the material balance is the production of 30 gal of ethanol/ton of dry WDF. Therefore, approach No. 2 above was used in adjusting Stream 22. Using this approach, the additional unconverted cellulose and the reduced amount of water affect Stream 24 only.

The yeast fermentation of glucose to ethanol is maintained at about 30 °C (86 °F) with chilled water. In geographic areas where cooling water is available at temperatures lower than the fermentation temperature, refrigeration systems (chillers) are not needed. Carbon dioxide can be purified and recovered through established technologies, the details of which are not included here.

The total steam required for the distillation plus dehydration step to obtain 200 proof ethanol (from 3.29 wt percent ethanol) is 3.33 lb/lb of ethanol produced. This value is consistent with the literature value for advanced distillation processes and has been used for the present balance.



Table 7

## Material Balance for a 400 ton/day MSW Plant (TVA)

Material	Quantities											
	Stream No.	1	2	3	4	5	6	7	8	9	10	11
Stream component	Quantity of MSW (TPD)	MSW to trommel (TPD)	Ferrous metals from trommel	Fines to stoner	Glass from stoner	Organic fines to hydrolysis	WDF to picking lines	HDPE to baler	PET to baler	LDPE to landfill	Nonferrous to storage	Paper to shredder
MSW	400.00	-	-	-	-	-	-	-	-	-	-	-
Paper and paperboard	-	123.90	-	-	-	-	123.90	-	-	-	-	74.34
Glass	-	39.60	-	39.60	33.66	5.94	-	-	-	-	-	-
Aluminum	-	5.60	-	-	-	-	5.60	-	-	-	-	-
Ferrous metals/tin/lead	-	35.40	26.55	-	-	-	8.85	-	-	-	-	-
Nonferrous metals	-	1.00	-	-	-	-	1.00	-	-	-	0.95	-
Plastics (HDPE)	-	10.80	-	-	-	-	10.80	10.26	-	-	-	-
Plastics (LDPE)	-	12.80	-	-	-	-	12.80	-	-	12.80	-	-
Plastics (PET)	-	5.60	-	-	-	-	5.60	-	5.32	-	-	-
Other plastics	-	11.20	-	-	-	-	11.20	-	-	-	-	-
Rubber and leather	-	11.22	-	-	-	-	11.22	-	-	-	-	-
Textiles	-	7.68	-	-	-	-	7.68	-	-	-	-	-
Food wastes	-	14.70	-	14.70	0.74	13.97	-	-	-	-	-	-
Wood	-	14.40	-	14.40	0.72	13.68	-	-	-	-	-	-
Yard wastes	-	10.20	-	10.20	0.51	9.69	-	-	-	-	-	-
Misc. inorganics	-	8.40	-	8.40	0.42	7.98	-	-	-	-	-	-
Water	-	87.50	-	28.92	0.91	28.00	58.58	-	-	-	-	28.34
Total	400.00	400.00	26.55	116.22	36.96	79.26	257.23	10.26	5.32	12.80	0.95	102.68

Table 7 (Cont'd)

Material	Quantities								
	Stream No.	12	13	14	15	16	17	18	19A
Stream component		Picking lines to aluminum separator	Aluminum to baler	WDF to magnet	Metals from magnet	Residue to air classifier	Heavies to land-fill	Lights to shredder	WDF to hydrolysis
MSW		--	--	--	--	--	--	--	--
Paper & paperboard		49.56	--	49.56	--	49.56	2.48	47.08	121.42
Glass		--	--	--	--	--	--	--	5.94
Aluminum		5.60	5.04	0.56	--	0.56	--	0.56	0.56
Ferrous metals/tin/steel		8.85	--	8.85	7.08	1.77	1.42	0.35	0.35
Nonferrous metals		0.05	--	0.05	0.05	--	--	--	--
Plastics (HDPE)		0.54	--	0.54	--	0.54	--	0.54	0.54
Plastics (LDPE)		--	--	--	--	--	--	--	--
Plastics (PET)		0.28	--	0.28	--	0.28	--	0.28	0.28
Other plastics		11.20	--	11.20	--	11.20	10.64	0.56	0.56
Rubber and leather		11.22	--	11.22	--	11.22	9.35	1.87	1.87
Textiles		7.68	--	7.68	--	7.68	5.44	2.24	2.24
Food wastes		--	--	--	--	--	--	--	13.97
Wood		--	--	--	--	--	--	--	13.68
Yard wastes		--	--	--	--	--	--	--	9.69
Misc. inorganics		--	--	--	--	--	--	--	7.98
Water		<u>30.24</u>	<u>--</u>	<u>30.24</u>	<u>--</u>	<u>30.24</u>	<u>3.11</u>	<u>27.14</u>	<u>83.48</u>
Total		125.22	5.04	120.18	7.13	113.05	32.43	80.63	262.57

Table 7 (Cont'd)

Material	Quantities												
	Stream No.	19B	20	21	22	23	24	25	26	27	28A	28B	29
Stream component		Biomass feedstock	Sulfuric acid solution	Steam	Acid biomass to press	Wash water	Solids to boiler	*	Hydrolyzate to furfural stripping	Furfural and acetic acid	Steam to stripping column	Condensate from stripping column	Hydrolyzate to neutralization
WDF		-	-	-	-	-	-	-	-	-	-	-	-
Hemi-cellulose		9.85	-	-	0.63	-	0.63	-	-	-	-	-	-
Cellulose		99.21	-	-	65.28	-	65.28	-	-	-	-	-	-
Lignin		27.76	-	-	27.76	-	27.76	-	-	-	-	-	-
Other		19.70	-	-	19.70	-	19.70	-	-	-	-	-	-
Ash		22.70	-	-	22.57	-	22.57	-	-	-	-	-	-
Xylose		-	-	-	4.92	-	0.15	-	4.78	-	-	-	4.78
Glucose		-	-	-	37.70	-	1.13	-	36.57	-	-	-	36.57
Furfural		-	-	-	2.95	-	0.09	-	2.87	2.87	-	-	-
Sulfuric acid		-	10.75	-	10.75	-	0.32	-	10.42	-	-	-	10.42
Water		83.48	321.07	121.97	522.75	136.30	129.26	-	529.79	-	48.42	48.42	529.79
Acetic acid		-	-	-	1.34	-	-	-	1.34	1.34	-	-	-
Carbon dioxide		-	-	-	-	-	-	-	-	-	-	-	-
Lime		-	-	-	-	-	-	-	-	-	-	-	-
Gypsum		-	-	-	-	-	-	-	-	-	-	-	-
Reheat		-	-	-	-	-	-	-	-	-	-	-	-
Total		262.56	331.82	121.97	716.35	136.30	266.89	-	585.76	4.21	48.42	48.42	581.56

\*Not Used

Table 7 (Cont'd)

Material	Quantities											
	Stream No.	30	31	32	33	34	35	36	37A	37B	38	39
Stream component		Lime	Hydrolyzate gypsum solution	Wash water	Gypsum to disposal	Hydrol. to ferment.	Carbon dioxide	Beer to distillation	Steam to distillation column	Condensate from distillation column	Sillage to evaporator	Ethanol to dehydration
WDF		-	-	-	-	-	-	-	-	-	-	-
Hemi-cellulose		-	-	-	-	-	-	-	-	-	-	-
Cellulose		-	-	-	-	-	-	-	-	-	-	-
Lignin		-	-	-	-	-	-	-	-	-	-	-
Other		-	-	-	-	-	-	-	-	-	-	-
Ash		-	-	-	-	-	-	-	-	-	-	-
Xylose		-	4.78	-	0.14	4.63	-	0.46	-	-	0.46	-
Glucose		-	36.57	-	1.10	35.47	-	3.55	-	-	3.55	-
Furfural		-	-	-	-	-	-	-	-	-	-	-
Sulfuric acid		-	-	-	-	-	-	-	-	-	-	-
Water		-	529.79	57.16	45.58	541.36	-	541.36	59.40	59.40	50.42	0.94
Acetic acid		-	-	-	-	-	-	-	-	-	-	-
Carbon dioxide		-	-	-	-	-	17.69	-	-	-	-	-
Lime		7.87	-	-	-	-	-	-	-	-	-	-
Gypsum		-	18.29	-	18.29	-	-	-	-	-	-	-
Ethanol		-	-	-	-	-	-	-	-	-	0.55	17.86
Total		7.87	589.43	57.16	65.11	581.47	17.69	563.78	59.40	59.40	544.99	18.80

Table 7 (Cont'd)

Material	Quantities						
	Stream No.	40	41	42	43	44A	44B
Stream component		Water from dehydration	Ethanol from dehydration	Makrop water to storage	Evaporator to silage boiler	Steam to silage processing	Condensate from silage processing
WDF		-	-	-	-	-	-
Hemi-cellulose		-	-	-	-	-	-
Cellulose		-	-	-	-	-	-
Lignin		-	-	-	-	-	-
Other		-	-	-	-	-	-
Ash		-	-	-	-	-	-
Xylose		-	-	-	0.46	-	-
Glucose		-	-	-	3.55	-	-
Furfural		-	-	-	-	-	-
Sulfuric acid		-	-	-	-	-	-
Water		0.94	-	537.75	2.67	180.64	180.64
Acetic acid		-	-	-	-	-	-
Carbon dioxide		-	-	-	-	-	-
Lime		-	-	-	-	-	-
Gypsum		-	-	-	-	-	-
Ethanol		-	-	0.55	-	-	-
Total		0.94	17.86	538.30	6.68	180.64	180.64

Also, 10.11 lb of steam/lb of ethanol produced goes to stillage processing, which is a significant amount. It is not clear if all the water present in Stream 38 entering the stillage processing must be evaporated and then condensed. If such is the case, then the steam requirement would increase significantly. The value has not been adjusted.

Overall, the plant produces 17.86 ton/day of 200 proof ethanol, or 5427.3 gal/day. This is equivalent to 13.57 gal/ton of as-fed MSW, or 20.67 gal/ton of as-fed WDF, or 30.31 gal/ton of dry WDF or 54.71 gal/ton of cellulosic material fed.

### *Enzymatic Hydrolysis (UAH)*

A material balance for a 400 ton/day MSW processing plant based on the UAH enzymatic hydrolysis process was also prepared. The material balance was based on scaling up the 27 ton/day case as described in the UAH study. Some substantial changes were made to the UAH flow sheet to render it comparable to that of the dilute acid hydrolysis case. Specifically, a front-end MSW classification system was added, identical to that in the TVA study. Also, a dehydration column was substituted for the pervaporation unit as originally specified in the UAH study. Further, instead of paper and waste paper, the feedstock to the scaled-up enzymatic hydrolysis process is WDF.

It was assumed that the composition of the paper and waste paper fed to the original UAH process was essentially 100 percent cellulose. Thus, for the initial stage of scale up, the cellulosic component of the WDF (Table 4, p 31) corresponds to the original UAH feed. The net result is that 262.56 ton/day of WDF is fed to the Steam Processor. The net material balance for the plant is summarized in Table 8; some details of the material balance are given in Table 9. Other steps and assumptions made to scale up the UAH process are described below. A revised process flow diagram of the enzymatic hydrolysis process is given in Figure 5.

In the hydrolysis section, WDF and recycled hot water are fed to the inclined rotating steam processors equipped with internal flights. The steam is added to the steam processors to raise the temperature to 290 °F. After depressurization in the flash tanks, over- and undersized inert materials, such as ash, are removed from the pretreated WDF in the vibrating screen.

The pretreated WDF is conveyed to a slurry tank and mixed with recycled enzyme/water solution and makeup enzyme. The slurry is then pumped to the hydrolysis reactors where the cellulosic fraction in the WDF is converted to fermentable sugars. The unreacted suspended solids are removed from the hydrolyzate through a solids recovery system, which consists of a centrifuge and a filter press. The solids are then dried and pelletized into solid fuel. The liquid hydrolyzate is pumped to an ultrafiltration unit to separate the enzymes for recycle. The resulting dilute sugar solution is further concentrated by reverse osmosis prior to fermentation. The water effluent from the reverse osmosis is recycled to the process. The fermentation, ethanol recovery, and utilities sections are similar to those for the dilute acid hydrolysis (TVA) process.

The capital costs were estimated based on major equipment costs for each processing section. Factors from Peters and Timmerhaus<sup>7</sup> for solid fluid processing plants were used to estimate the total capital costs. The first step was to eliminate the complication of adding 5 ton/day of hammer-milled paper downstream from the Steam Processor. This material was subtracted from the overall material balance. Next, the material balance was revised to account for additional species (hemicellulose, lignin, ash etc.) in the feedstock. The hemicellulose, lignin, ash, and others are not converted in the process.

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<sup>7</sup> Max S. Peters and Klaus D. Timmerhaus, *Plant Design and Economics for Chemical Engineers* (McGraw-Hill, 1980).

Table 8

**Net Material Balance for Enzymatic Hydrolysis Process  
(Based on 400 ton/day MSW or 262.56 ton/day WDF Feed)**

Material	Quantity (tons)
<b>Inputs</b>	
Waste derived feedstock (WDF)	262.56
Steam	338.27
Water	207.61
Enzyme	0.47
Yeast	0.14
Ammonia	<u>2.77</u>
Total	811.82
<b>Outputs</b>	
200-proof ethanol product	19.37
Solid fuel product	106.43
Fusel oil product	1.05
Yeast product (70% moisture)	3.28
Carbon dioxide product	19.64
Vibratory screen waste (>1/2")	42.01
Condensate	485.98
Evaporated Moisture	<u>134.06</u>
Total	811.82

The ash and others are removed by the 2-in. and 1/2-in. screens (Stream 9), while the hemicellulose and lignin are removed with the solid fuel (Stream 26).

The quantity of steam feed to the "Steam Processor" was estimated by energy balance. Based on the UAH literature, it was assumed that one-third of the steam feed is condensed in this step, while two-thirds escapes as steam during depressurization. The condensed steam becomes part of the solids leaving the steam processor (Stream 4). The steam required was estimated to be 0.526 lb/lb of as-fed WDF or 0.77 lb/lb of dry WDF. The amount of hot water entering the Steam Processor (Stream 3) is 0.299 lb/lb of as-fed WDF or 0.438 lb/lb of dry WDF. The steam and hot water feed rates are based on 80 °F WDF, 160 °F hot water, and 293 °F steam (60 psia, saturated).

The specific heat of moisture-free WDF was estimated by the following formula:

$$C_p = 0.266 + (0.00116 \times T)$$

where  $C_p$  = specific heat, kcal/g-°C, and  $T$  = temperature, °C. This formula is based on wood on an oven dry basis.<sup>8</sup>

The solids leaving the Steam Processor (Stream 4) contain 60.28 percent moisture (as in the UAH design), based on its cellulose, hemicellulose, and lignin contents only (i.e., ash and others fractions are not considered).

The concentration of cellulose entering the Hydrolysis Tank (Stream 14) is 2.32 percent (as in the UAH design). The total solids concentration is higher due to the noncellulosic portion of the feed.

<sup>8</sup> H.L. Chum, L.J. Douglas, D.A. Feinberg, and H.A. Schroeder, "Evaluation of Pretreatments of Biomass for Enzymatic Hydrolysis of Cellulose," SERI/TR-231-2183 (October 1985), p 77.

Table 9

**Material Balance for a 400 ton/day MSW or 262.56 ton/day WDF Plant**  
**Based on the Enzymatic Hydrolysis Process**

Stream No.	1	2	3	4,8	5	9	10,11	14	15,17	22	26	33
Description	WDF Feed	Steam	Hot Water	Feed to Screens	Vented Steam	>1/2 in. Waste	Feed to Hydrol. Reactors	Str. 11+ Recycled H <sub>2</sub> O+Enz	Slurry to Centrifuge	Solids to Fuel Dryer	Solid Fuel (10%)	Glucose to Ferment.
Component												
Cellulose	99.25	-	-	99.25	-	-	99.25	99.25	57.98	57.98	57.98	-
Hemicellulose	9.98	-	-	9.98	-	-	9.98	9.98	9.98	9.98	9.98	-
Lignin	27.83	-	-	27.83	-	-	27.83	27.83	27.83	27.83	27.83	-
Ash	22.58	-	-	22.58	-	22.58	-	-	-	-	-	-
Others	19.43	-	-	19.43	-	19.43	-	-	-	-	-	-
Water	83.49	-	78.43	207.98	-	-	207.98	3695.41	3690.82	143.68	10.64	231.36
Steam	0.00	138.17	-	-	92.11	-	-	-	-	-	-	-
Glucose	-	-	-	-	-	-	-	-	45.85	-	-	45.85
Total	262.56	138.17	78.43	387.05	92.11	42.01	345.04	3832.46	3832.46	239.47	106.43	277.21





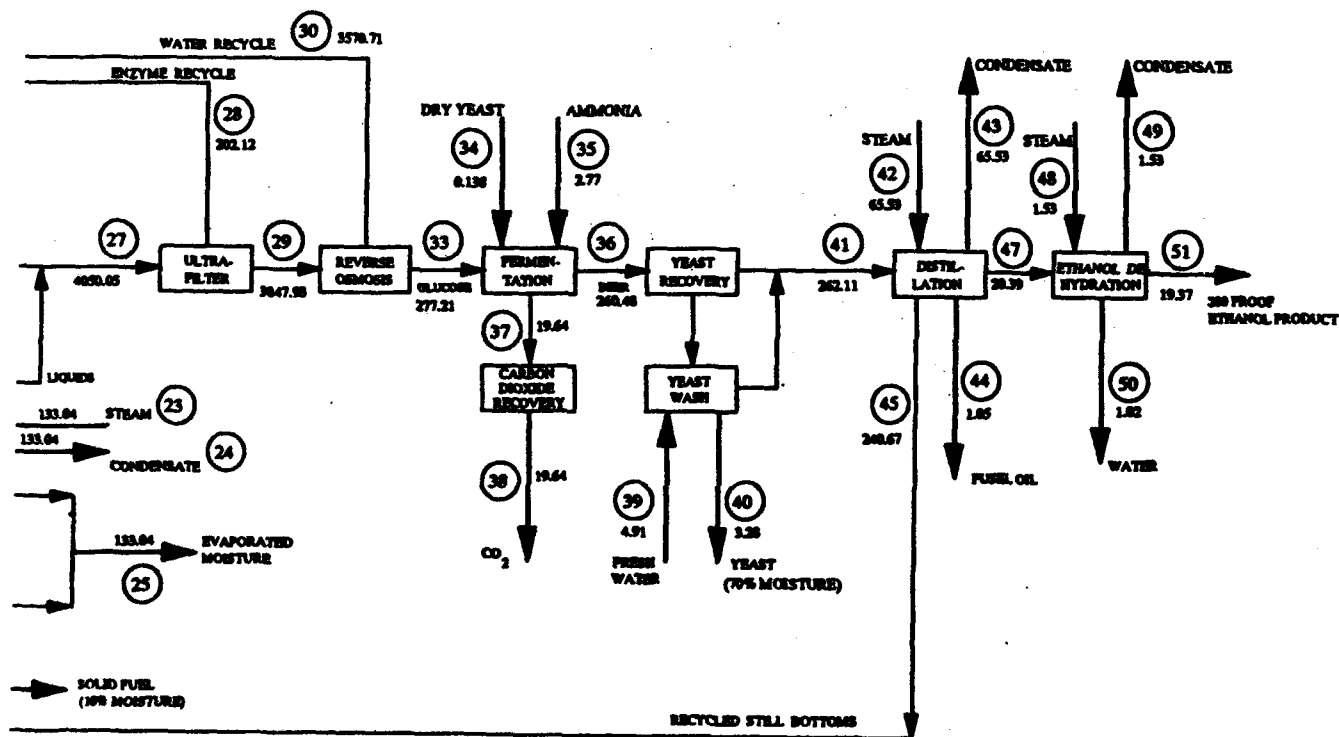


Figure 5. (Cont'd).

The moisture content of the solids entering the Fuel Dryer (Stream 22) is 60.0 wt percent (as in the UAH design). The pelletized Solid Fuel (Stream 26) contains 10 percent moisture (based on UAH literature). The fraction of cellulose leaving in this solid fuel stream is the same as in the UAH design.

The wash water required/pound of dry solids recovered (Stream 21) is 4.77 (as in the UAH balance). A significant amount of additional wash water is required due to the presence of hemicellulose and lignin in the feed. The amount of steam required to evaporate this additional moisture (Stream 23) is significant.

A portion of the enzyme is lost in the hydrolysis step due to absorption onto the unhydrolyzed fraction of the cellulose. Since the feed now contains noncellulosic material, some additional enzyme will be lost from the system; however, no allowance has been provided for it.

The Recycled Still Bottom (Stream 45) is expected to contain some solids that must be purged from the system. However, the UAH balance has ignored this processing aspect; it is not considered.

In the UAH balance, the total steam required for the distillation step to obtain 200 proof ethanol (from 7.39 wt percent ethanol) is 3.46 lb/lb of ethanol produced. This amount is consistent with the literature value for advanced distillation processes and has been retained in the present balance. As mentioned above, a dehydration column was substituted for the pervaporation unit; however, the total steam requirement should be similar. (It should be noted that pervaporation would not require the use of a third component, such as benzene or hexane to affect the separation of the water and ethanol and by virtue of this, may be more environmentally acceptable.)

The steam vented from the Steam Processor goes to Hot Water Storage, where it is condensed while heating the water. Since the temperature of the water entering and leaving Hot Water Storage is unknown, it is problematic to make an energy balance on this subsystem. However, it appears that the heat available in the vented steam exceeds that required to heat up the incoming water, thus this atmospheric pressure steam may be used elsewhere, if possible.

The enzyme and ethanol fermentations are maintained at about 30 °C (86 °F) with chilled water. In areas where the cooling water is available at a temperature sufficiently lower than the fermentation temperature or if thermophilic microorganisms can be used in the enzyme production at a temperature sufficiently higher than the cooling water temperature, refrigeration systems (chillers) are not needed.

The carbon dioxide can be purified and recovered through established technologies, although the details of this step have not been considered here.

The plant produces 19.37 ton/day of 200 proof ethanol or 5886.2 gal/day. This ethanol production is equivalent to 14.71 gal/ton of as fed MSW, or 22.42 gal/ton of as fed WDF, or 32.87 gal/ton of dry WDF, or 59.31 gal/ton of cellulosic material feed.

## 4 MSW UTILIZATION ALTERNATIVES

### Introduction

Municipal solid waste (MSW) can be used or disposed of in several ways. The most common form of MSW use at present is by landfill. However, landfilling has become an enormous problem; over 200 million tons of garbage are discarded each year, and this rate is increasing at an annual rate of 4 percent.<sup>9</sup> According to the USEPA, more than half of the existing landfills in the United States will reach their capacity within the next 8 years. A number of alternatives to MSW landfilling are available that allow for some productive use of the waste. One of the most useful applications of MSW is as an energy source to provide steam for industrial applications or to generate electricity for utilities or cogenerators. Based on an average calorific value of 4500 Btu/lb, MSW represents an available energy source of nearly 1.8 quads ( $10^{12}$  Btu)/year. This is the equivalent of nearly 60 million tons of coal, or about 10 percent of the annual electric utility consumption.<sup>10</sup> However, the alternative applications of MSW, of which there are several, should not be considered to be mutually exclusive. For example, many experts now believe that a balanced program of landfilling, recycling, and incineration with energy production will ultimately be adopted to address the MSW problem. The following sections provide an overview of MSW utilization alternatives available at the present time, including a brief description of the technology, a discussion of the major manufacturers and existing facility operators, and a preliminary estimation of costs.

### Landfilling

Landfilling has been the practical solution for disposal of MSW for many years, and will remain an important component of the solid waste industry for the foreseeable future. At present, about 83 percent of MSW is disposed of in sanitary landfills, many of which are becoming full or are experiencing environmental compliance problems. Major environmental problems include, but are not limited to, restrictions on landfill gas and contamination of fresh groundwater supplies. As a result of these environmental and capacity constraints, nearly three-quarters of existing landfills will close within the next 15 years. Recent estimates have indicated that by the year 2000 the United States will be short of MSW disposal capacity by 56 million ton/year. Obviously, serious steps need to be taken to reduce the quantity of wastes flowing into remaining landfills. Materials recovery through recycling will play a very important role in reducing the quantity of MSW, but recent experience has shown the existence of a practical limit of slightly over 30 percent of the waste stream that can be economically recycled given public participation and the market for recycled products.

The cost of landfilling, often called a "tipping fee," varies widely across the nation and often varies significantly even within a given region. Tipping fees at landfills currently average \$45 to \$55/ton, and have been increasing at double digit rates in recent years. Tipping fees will play an important role in the economics of alternative MSW utilization technologies. For example, tipping fees collected by MSW mass burn operators typically provide 25 to 100 percent of the revenue collected, with the remainder coming from sales of steam or electric energy and, for some facilities, sale of recycled materials collected from the waste stream. Thus, the cost of landfilling MSW will play an important role in determining the cost effectiveness of alternative MSW utilization technologies.

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<sup>9</sup> Daugherty, Brooks, and Read, "Unloading Our Landfills," *Environmental Protection* (July/August 1991), pp 26-29.

<sup>10</sup> *Guidelines for Cofiring Refuse-Derived Fuel in Electric Utility Boilers, Volume 1: Executive Summary* (Electric Power Research Institute [EPRI], June 1988).

## Recycling

Municipal solid waste may contain significant portions of recyclable material. Although recycling will not eliminate all solid wastes (and thus does not represent a disposal option), an effective large scale recycling program can reduce the quantity of solid waste generated by as much as one-third. Generally, recycling can take place before the material enters the waste stream (source separation) or before being processed as waste (end use classification). Many communities are now mandating recycling programs that will remove the majority of recyclable materials from MSW. Thus, the composition of MSW is expected to change as recycling programs proliferate. The ability to extract recyclable materials from the waste stream is an important consideration in establishing certain types of waste-processing facilities such as refuse-derived fuel (RDF) facilities, because these facilities typically count on revenue from sale of recyclables to offset a portion of facility operating costs.

The cost of MSW disposal through the use of a combination of recycling and landfilling of the remainder is dependent on landfill costs, local markets for recyclables, and material collection techniques (mixed waste processing versus source separation). The trend for local markets can be predicted to a certain extent since high landfill costs drive solid waste generators to recycling and produce mandated recycling. Extensive local recycling then tends to glut the market for certain materials. State laws requiring a minimum recycled content in certain products promise to reduce the glut.

Disposal costs at facilities with a recycling "front end" range from \$9/ton to \$92/ton.<sup>11</sup> For comparative purposes in this analysis, a 400 ton/day material recovery facility in the midwest using a combination of source separation and mixed waste processing would have a disposal cost of \$31/ton.

One important factor to recognize is that recyclable materials usually face a "buyers market." Although subject to the forces of supply and demand like all commodities, recycle market prices are generally influenced more heavily by buyers than by sellers. Local recycle markets are often dominated by a few large buyers for each type of material. Due to the relatively low value of the recyclable materials (with the possible exception of aluminum), recyclables cannot overcome the high cost of transportation to distant markets. Thus, recycle markets are usually local. The consensus opinion for recyclable materials is that future markets (and prices) should improve for aluminum, plastic, and corrugated containers. Recycling of newsprint will continue to expand but prices will remain low. Markets for steel and glass should be stable.<sup>12</sup> The following discussion provides an overview of the markets for various recyclable materials, along with the prices that are occurring at the present time.

### *Paper and Newsprint*

Paper products comprise the largest component of MSW. About 40 percent by weight of MSW is represented by paper and paperboard products. Old newspaper (ONP) alone accounts for 19 percent of total solid waste. For each ton of wastepaper recycled, 3 cu yd of landfill space are saved. According to the American Paper Institute (API), 28.9 million tons of wastepaper were collected in 1990. About 77 percent (22.4 million tons) was shipped to U.S. mills for making recycled paper, with the remaining 22 percent (6.5 million tons) exported. In fact, the United States is by far the world's largest exporter of wastepaper. The collection rate for old newspaper approaches 45 percent of the 13.3 million tons of newsprint consumed in the United States, up 9 percent from 1989. API estimates that recycled ONP had the following end uses: 40 percent as newsprint, 37 percent as folding cartons, 6 percent as sanitary tissue, and 17 percent for other uses such as construction paper and board, molded egg cartons, cushioning

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<sup>11</sup> 1990 Resource Recovery Yearbook.

<sup>12</sup> Solid Waste Management Newsletter (Office of Technology Transfer, University of Illinois Center for Solid Waste Management and Research, June and July 1991), Vol 5, No. 6 and 7.

materials for packaging, animal bedding, and insulation.<sup>13</sup> Paper types other than ONP that can be recycled include high grade paper such as office and copier paper, old corrugated containers (covered in the next section), pulp substitutes such as brown kraft and semibleached sheets, and mixed papers of varied quality. Of these types, mixed papers carry the lowest market value (sometimes even zero) due to the need to separate the paper types before recycling. ONP prices are highly volatile and vary considerably among locations. Prices for old newspaper, the most actively traded and recycled type of paper, can range between \$15 and \$60/ton. High grades tend to experience less volatile fluctuations, with prices in the range of \$150 to \$250/ton. These higher quality papers require less pulping and can be used as a feedstock for a larger number of paper products, and thus have greater value. In general, wastepaper prices react quickly to changes in the pulp and paper industry, and are considered a bellwether of future market conditions. Many wastepaper recycling mills are presently running near full capacity, and enjoy an oversupply of wastepaper. The direction of prices is generally determined by the demand for recycled paper from consumers. Most experts predict continued sharp price swings in wastepaper prices.

To stimulate demand for recycled paper, a number of states and municipalities have imposed a tax on newspapers that do not use some portion of recycled paper content. The U.S. government, through the 1989 Resource Conservation and Recovery Act, has mandated that all government agencies must implement a preference purchasing program favoring recycled items. For example, newsprint must contain 40 percent wastepaper; duplicating paper and stationery, 50 percent; toilet tissue, 20 percent; and corrugated cartons, 35 percent.

Significant investment in repulping and de-inking equipment is occurring in North American paper mills. Nine Canadian and seven U.S. recycled newsprint mills are expected to be completed during 1991 and 1992. Among the companies conducting these expansions are Stone Container (in a joint venture with Waste Management), Paper Recycling International, Browning-Ferris Industries (in a joint venture with Weyerhaeuser), and Jefferson Smurfit. These facilities are expected to increase recycled newsprint capacity from 2.4 to 4.2 million metric tons. However, 56 percent of this capacity will be in Canada, although the mills may draw significant supplies from U.S. markets in the Northeast and upper Midwest. An additional 18 recycled newsprint mills are under consideration. Some newsprint and corrugated recyclers, including Stone Container and Green Bay Packaging, do not de-ink their product. Rather, they sell corrugated containerboard or kraft paperboard that is darker or somewhat speckled in appearance.

Despite the increased activity in paper recycling, solid waste industry experts are anticipating that 11 to 13 million tons of ONP will still be disposed of by other means in the year 2000.<sup>14</sup> In general, the recycled paper market is expected to suffer from oversupply despite the presence of major new recycling facilities in the next few years. Mandatory collection laws are contributing to this trend. These factors have important implications for MSW use technologies that benefit from the presence of wastepaper in the waste stream.

### *Corrugated Containers*

Old corrugated containers (OCC) account for 13 percent by weight of MSW. Most of the recycled OCC is collected from retail stores, grocery stores, factories and office buildings. Less than 25 percent of OCC is collected from consumer uses. Corrugated cardboard is widely used for shipping containers, and thus tends to accumulate at businesses. OCC is recycled at mills to make both the linerboard as well as the corrugating medium for new corrugated cardboard boxes, cereal boxes, shoe boxes and cores for paper roles. API reports that OCC reached a recycling rate of 50 percent in 1990 (13.2 million tons). Of the total collected, 10.4 million tons were recycled at domestic mills and 2.7 million tons were

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<sup>13</sup> *Pulp & Paper 1990 North American Fact Book* (Miller Freeman Publications, San Francisco, CA, 1990).

<sup>14</sup> J. Hastreiter, "Paper Recycling Industry Shifts Into High Gear," *World Wastes* (July 1991), pp 42-48.

exported. About 80 percent of collections were used containers, while the remaining 20 percent were container plant cuttings, which are slightly more valuable than old containers. The paper industry estimates that present recycling technology will limit the recycling level of OCC to 66 percent of total OCC generation. Supplies of OCC are sufficient at the present time, but will be constrained if the recycling rate does not rise above 60 percent by 1995. Consumer recycling programs will need to be implemented if the 60 percent recycle goal for OCC is to be met. Demand for OCC, mainly from linerboard and corrugating medium mills, has risen dramatically in recent years. This caused prices to double in many markets. Some industry analysts believe that prospects for future shortages of OCC, as well as the demand created by several new corrugated mills now being built, will cause a continued increase in the price of OCC. Current prices for OCC are in the range of \$40 to \$70/ton, while double-lined kraft corrugated clippings are in the range of \$70 to \$125/ton. Companies with new OCC recycling plants under construction include Gaylord Container and several by Stone Container.

### *Aluminum*

Aluminum represents about 2 percent by weight of the MSW stream, but contributes about 40 percent of the material revenues to recycling programs. About 60 percent of aluminum wastes are cans, and these cans were recycled at a rate of over 50 percent in 1990. Aluminum is typically sold directly to primary aluminum producers. Typical aluminum UBC (used beverage container) prices are in the range of \$0.45 to \$0.65/lb. However, recent evidence indicates that the market for recycled aluminum is becoming more volatile and increasingly segmented. Previously, the major aluminum companies (Alcoa, Reynolds, American National Can) tended to act in unison. Now, smaller aluminum producers and secondary aluminum smelters are occupying a larger market position and are paying widely varying prices (often higher) for aluminum.

### *Ferrous Metals*

Ferrous metal scrap represents about 7 percent by weight of the MSW stream (excluding automobiles). The primary form of ferrous metals recovered in recycling programs are tin-plated steel cans, which represent about 2 percent of the MSW stream. In the past, the primary buyers of these cans were detinning mills, which separated the tin component from the high grade steel. A recent shift has seen the entry of steel mills into the market for tin plated steel cans, due to new technology that allows much higher amounts of tin into the steel melt than previously (5 to 10 percent tin rather than 1 percent). However, steel mills generally have stricter requirements on cleanliness and baling than do detinning mills. In addition, joint ventures between detinning mills and steel makers are coming into existence to provide quality control for incoming scrap. Prices for recycled ferrous metals are generally \$40 to \$60/ton from detinners and \$60 to \$80/ton from steel makers.

### *Glass*

Glass comprises about 7 percent by weight of the MSW stream. Most of the recycled glass is derived from beverage bottles and food jars, which are free of the impurities that exist in other glass products (e.g., cookware, mirrors, light bulbs, window glass). The main market for recycled glass is as cullet for making new glass containers, which provides a cost savings to glass manufacturers because the melting of cullet requires lower operating temperatures than the melting of sand. Less than 10 percent of recycled glass is used in secondary products such as insulation. Quality is very important in glass recycle markets because contaminants can create problems when the cullet is melted. In addition, recycled glass is considerably more valuable if it can be sorted according to color. The current market price for recycled glass is \$30 to \$70/ton, depending upon color and quality. Mixed-color glass cullet brings much lower prices in the range of \$10 to \$30/ton.

## *Plastics*

Plastics account for 8 percent by weight of the MSW stream, but represent 18 percent by volume. Plastic use, and thus its proportion of solid waste, has increased dramatically over the past 20 years. The major constraint to plastic recycling is the different types of plastics in common usage, differentiated by the number of resins used in their manufacture. These resins create problems in the recycling process because they have different melting points and molecular structures that do not bond well together. To alleviate this problem, the Society of the Plastics Industry has developed a coding system whereby an identifying number and symbol are molded into each container. The plastic coding system is identified in Table 10.

Most recycling programs focus on PET and HDPE, which have ready markets and are commonly found in large quantities in waste streams. Recycle markets are rather limited for the other plastic types. HDPE and PET are often sold to processors who grind the plastic into chips, wash them, and sell them to a plastic product manufacturer. Most of the plastic resin brokers and scrap handlers who perform the processing are smaller companies, although large companies (such as DuPont, Mobile and Dow Chemical) are becoming increasingly involved and are advancing plastic recycling technology. Presently, the growth of plastic recycling is supply-constrained. Recent prices for recycled plastics were \$120 to \$200/ton for PET and \$80 to \$150/ton for HDPE.

## *Composting*

An additional technology frequently used in conjunction with recycling is composting. The extent of this technique can range from simply composting the municipalities yard/landscaping waste, to composting the entire biodegradable portion of the waste stream. MSW composting is a technology that competes directly with incineration for the organic component of the waste stream. The compost produced at these facilities is typically given away for landscaping purposes and should not be thought of as a marketable commodity.<sup>15</sup> The average tipping fee at recycling/composting facilities with a throughput of approximately 400 ton/day is \$32/ton.<sup>16</sup>

## *Mass Burning*

Mass burning of MSW involves the combustion of the waste stream with very little, if any, removal of noncombustibles. Mass burn plants charge a tipping fee for accepting the MSW. Energy is usually, but not always, recovered in the form of steam or electricity, which provides an additional form of revenue to the mass burn operator. The first large scale mass burn plants in North America were built in the early 1970s, with significant technological advances since then in the areas of combustion efficiency, increased steam pressures, computerized plant controls, and advanced emissions controls. At many mass-burn waste to energy (WTE) plants, availability now equals that of utility grade power plants. Mass-burn facilities are available as modular, shop-assembled units, or are field-erected at the site. Large-scale mass-burn plants (greater than 500 ton/day) are field-erected, whereas smaller scale plants are a mixture of field-erected and shop-fabricated. Mass burn plants report an average material reduction of 70 percent by weight and an average use of 85 percent of rated capacity. MSW incineration involves drying, devolatilization, and ignition of waste. Controlled air combustion and excess air combustion are the two most prevalent types. Controlled air combustion isolates the ignition step by combusting volatilized gases in a chamber separate from where the MSW is dried, volatilized, and turned to char. Excess air combustion completes drying, devolatilization, and ignition in the same chamber. Excess air combustion

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<sup>15</sup> T. Watson, "Solid Waste Composting Aims for the Mainstream," *Resource Recycling* (July 1990), pp 30-40.

<sup>16</sup> S. Apotheke, "Garbage In, But What Comes Out?", *Resource Recycling* (September 1990), pp 40-45.



Table 10

## Plastic Identification Coding System

No.	Abb.	Type	Usage
1	PET	Polyethylene terephthalate	Soda bottles
2	HDPE	High density polyethylene	Water and milk jugs, oil bottles
3	PVC	Polyvinyl chloride	Pipe, flooring, some food packages
4	LDPE	Low density polyethylene	Plastic bags, wire insulation
5	PP	Polypropylene	Battery casings, bread wrap
6	PS	Polystyrene	Foam food containers
7	--	Other	Multilayered packaging

is also typically used in Refuse Derived Fuel facilities. Mass burning of unprocessed waste typically employs either refractory-lined or water wall furnaces. In refractory-lined furnaces, heat is recovered downstream of the furnace using a waste heat boiler. The furnace maintains elevated combustion temperatures and can accommodate very low Btu fuel. In water wall furnaces, both the water wall and convection (back) pass tubes are used for heat recovery and steam generation. The direct radiative and convective heat transfer provided by the fireball in the furnace results in a heat recovery efficiency greater than that of the refractory type, which is primarily convective heat transfer. Another advantage of the water wall furnace is lower excess air requirements due to the lower furnace temperature provided by the boiler tubes, resulting in the potential for lower capital costs related to moving and processing the combustion air. Mass burning can also be done in modular starved air combustors, consisting of a refractory-lined furnace with a deficient supply of combustion air, a second stage chamber to complete the burning, and a heat recovery boiler for generating steam.

Currently, there are 140 operating mass burn projects in the United States with a total capacity of 21.1 million ton/day, and 62 plants are under construction or in advanced developmental stages with capacity of 26.8 million ton/day.<sup>17</sup> A total of 94 projects in conceptual and advanced planning stages are scheduled to become operational in the next 5 years. At present, 18 percent of all MSW is being disposed of in mass-burn WTE plants. When all 62 plants now under construction or in advanced development are completed, mass-burn WTE plants will consume about 28 percent of all MSW. The average tipping fees at existing plants have increased by 47 percent in the past 3 years and by 161 percent since 1986. Although tipping fees vary widely across the nation (from \$3 to \$270/ton), the average nationwide tipping fee at mass-burn WTE facilities is \$45.34/ton.<sup>18</sup>

A trend towards very large plants is now underway; in the 3 years between 1988 and 1990, average plant capacity nearly doubled from 574 to 1056 ton/day.<sup>19</sup> The increase in plant size has also brought about a greater role for electric power generation as a significant component of WTE economics. Only 58 percent of the plants operating before 1988 generated electricity, while 88 percent of the plants operating or under construction between 1988 and 1990 will generate electricity. The average energy generating capacity/plant has increased from 19.6 to 31.7 MW during the same time frame. Plants currently in advanced planning stages show an average generating capacity of 33.2 MW. A good indication of recent activity in the mass burn industry is that 63 percent of all existing plants have come on line since 1985.

<sup>17</sup> 1990 Resource Recovery Yearbook.

<sup>18</sup> S. Darcey, "WTE Recovers From Industry Shakedown," *World Wastes* (June 1991), pp 30-45.

<sup>19</sup> Hocker, p 37.

### *Mass Burn Ownership and Operation*

The mass burn WTE industry has shifted from predominantly public ownership and operation to a much greater private sector involvement; more than half of the plants now under construction will be privately owned, and more than 90 percent will be privately operated. Among presently operating plants, 60 percent are publicly owned, while about 40 percent are publicly operated. Numerous partnerships have been developed between municipalities and private developers to control costs and transfer the risks of development and operation. Private, full-service WTE companies have become increasingly involved in the industry, offering complete development, engineering, construction, and operation services to municipalities and other public agencies. Table 11 provides a list of the full service vendors that are involved in the mass burn WTE industry, along with the number and capacity (in ton/day) of plants that each vendor has in service, under construction, and awaiting a permit.

### *Mass Burn Economics*

A survey was conducted in 1990 of 17 field-erected mass-burn facilities with energy recovery.<sup>20</sup> The surveyed facilities had no more than 550 ton/day of capacity, with an average capacity of 333 ton/day. The survey reported that the average capital cost of these facilities was \$86,815/ton/day. The capital cost of these facilities ranged from \$55,000 to \$141,900/ton/day of capacity. All costs in the survey were adjusted to 1987 dollars for comparative purposes. Assuming that mass-burn capital costs have risen at the same rate as general inflation, then 1991 capital costs should be about 14 percent higher than in 1987, which makes the 1991 average capital cost slightly less than \$100,000/ton/day of capacity. The report also surveyed annual operating costs at those same field-erected mass burn facilities below 550 ton/day with energy recovery. The report showed that annual operating costs amounted to an average of \$13,333/ton/day of capacity. Again, this cost was reported in 1987 dollars, so the annual operating cost for 1991 is expected to be slightly over \$15,000/ton/day.

### *Mass Burn Environmental Compliance*

As mandated by the Clean Air Act Amendments of 1990, the USEPA has promulgated strict new source performance standards (NSPS) for air emissions from municipal waste combustors and RDF combustors with a capacity greater than 250 ton/day (on a per facility, not per unit, basis). Not covered by the NSPS are cofired combustors that burn fuel feed stream that is less than 30 percent comprised of MSW or RDF. Under the new EPA standards for municipal waste combustors, plants that began construction after December, 1989 must reduce heavy metals emissions and organic emissions of dioxins and furans by more than 99 percent by 1994. Also by 1994, sulfur dioxide and hydrogen chloride must be reduced by 90 and 95 percent, respectively. Nitrogen oxide emissions must be reduced by 40 percent. Those facilities in place before the 1989 cutoff date must be retrofitted to reduce heavy metal emissions by 97 percent, organic emissions by 95 percent, and acid gases (SO<sub>2</sub>, HCl, NO<sub>x</sub>) by 75 percent. The standards also include the requirement that facilities follow Good Combustion Practices (GCP), and specify a carbon monoxide (CO) level to be attained for different types of combustion technologies. The limits on CO concentrations are 50 ppm<sub>dv</sub> for modular starved air and excess air systems, 100 ppm<sub>dv</sub> for fluidized bed and mass-burn refractory wall/water wall systems, and 150 ppm<sub>dv</sub> for RDF, RDF/coal, and rotary water wall combustors. These are 4 hour average values corrected to 7 percent O<sub>2</sub>. New standards are also placed on particulate matter emissions.

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<sup>20</sup> B.A. Hegberg, W.H. Hallenbeck, and G.R. Brenniman, *Municipal Solid Waste Incineration With Energy Recovery (Technologies, Facilities and Vendors for Less Than 550 Tons Per Day)*, Report No. OTT-4 (University of Illinois Office of Technology Transfer, March 1990).

Table 11

## Full-Service, Mass-Burn WTE Industry Vendors

Primary Vendor	Operating	Construction	In Permitting
Ogden Martin	20,225 (20)	750 (1)	8,260 (6)
Wheelabrator	18,850 (20)	3,050 (2)	6,750 (3)
ABB	7,500 (3)	--	3,050 (2)
Montenay Power	7,200 (7)	1,250 (2)	1,500 (1)
American Ref-Fuel	5,010 (2)	620 (1)	5,720 (4)
Foster Wheeler	4,410 (5)	400 (1)	3,440 (3)
Consumat	4,196 (29)	--	--
Westinghouse	2,254 (3)	2,688 (1)	5,140 (4)

## Refuse Derived Fuel

RDF is made from MSW via a sequence of steps to reduce particle size and remove noncombustible materials, especially those materials that have value as recyclables. RDF facilities are generally very labor and energy intensive on the "front end" processing steps that classify, separate, shred, and screen the wastes. After being processed into a homogenous combustible fuel, RDF is often compacted into dense pellets or slugs to make fuel handling more convenient. However, RDF can also be used in raw, coarse, fluff, and powder forms. Densified RDF has an energy value of 7500 Btu/lb, similar to high grade lignite coal, and can be easily used in power plants, cement kilns, and cofired with coal-burning facilities. Nondensified RDF has a Btu value of 4500 to 6000 Btu, depending upon the type of preparation. Processing of wastes into RDF results in a uniformly sized product with a higher energy content than standard MSW that can be handled more easily and fired in more conventional boiler systems. The removal of noncombustibles raises the performance of the boiler and reduces slagging and jamming of combustor grates in comparison to standard MSW. In terms of the potential for SO<sub>2</sub> emissions during combustion, RDF has lower sulfur content (0.2 percent) than low sulfur bituminous coal (0.7 percent). Although it can be combusted without the aid of other fuels, RDF is often cofired with coal in existing coal-fired heat plants. The incentive for using cofiring RDF is a reduction in coal consumption of 5 to 15 percent without a significant deterioration in rated capacity. Best of all, RDF cofiring does not require major new investment or extensive modification to existing coal-firing equipment.

With respect to the preceding discussion on mass-burn technology, it appears that the differences between mass-burn and RDF may be considerably less important in the future due to the growing effects of recycling and resource recovery programs. These programs remove a large portion of noncombustible materials from the waste stream prior to incineration, thus making the solid waste entering a mass-burn facility very similar to processed RDF. This appears to be beneficial for mass-burn plants; the Foster Wheeler Company found that a local recycling program in Camden, NJ resulted in a 10 percent increase in the fuel heating value due to more complete combustion of MSW.

At present, 51 WTE facilities in the United States conduct recycling and materials recovery as an integral part of current operations. Since 1970, nine electric utilities have cofired RDF with coal in electric utility boilers. Of these, 4 continue to cofire RDF while the others have discontinued cofiring for various technical and economic reasons. One of the major firms involved in RDF plants is Asea Brown Boveri (ABB), which recently brought three large RDF plants on line in Honolulu, Detroit, and Hartford, CT. These facilities have a total capacity of 7500 ton/day. ABB is also developing a 2250 ton/day RDF facility in Chester, PA.

## *RDF Economics*

The value of RDF is not equal to the replacement value of coal. Energy produced from RDF is less valuable than energy produced from coal because the net fuel savings resulting from reducing coal consumption are at least partially offset by the incremental capital, operating and maintenance costs of RDF cofiring. Thus, a long-term, site-specific analysis of coal costs and facility requirements should be performed before a commitment to RDF is made. A facility that processes RDF from MSW must consider additional factors. The revenue received from sale of recyclable materials must be carefully evaluated, taking into consideration the prospect of future prices for items such as aluminum, glass, ferrous metals, and wastepaper. In addition, the effect of recycling programs upon the waste stream must be considered if such programs do not already exist. Finally, the capital cost of the RDF processing equipment must be considered.

To provide a simple example, consider a 500 ton/day RDF plant that processes MSW into RDF fuel and produces electricity. Assume that the facility can process RDF with a net cost of \$10/ton; the net cost indicates the cost after credits have been taken for sale of recyclable materials and for collection of tipping fees. The \$10/ton cost includes all incremental operating and maintenance costs as well as capital recovery costs for the facility. The cost of RDF at \$10/ton is equivalent to coal at a price of \$0.85/million Btu. If the cost of coal is \$2.00/million Btu, then the annual savings to the facility from using RDF is approximately \$2.3 million, computed in the following manner:

$$(\$2.00 - \$0.85)/10^6 \text{ Btu} \times 6000 \text{ Btu/lb} \times 2000 \text{ lb/ton} \times 500 \text{ ton/day} \times 340 \text{ day/year}$$

Note that the value of electricity is not an essential ingredient of this analysis; the critical factor is the relative cost of RDF and coal. Given that electricity is priced high enough to justify its generation, then the least expensive alternative between coal and RDF should be adopted. This analysis framework is valid whether RDF is cofired or used as a single fuel. However, the critical number to be established is the net cost of processing RDF. The net cost includes sales of recyclable items as well as tipping fees, which are both subject to a considerable amount of uncertainty. In addition, the incremental operating and maintenance costs attributable to RDF are also difficult to establish. The average tipping fee at 400 ton/day RDF facilities is \$40/ton.<sup>21</sup>

### *Effects of RDF on Heat Plant Equipment*

A number of adverse effects on coal-firing equipment are created by cofiring RDF. Increased operation and maintenance costs will be incurred for the fuel-handling equipment. Increased furnace slagging and fouling will occur because RDF ash exhibits poor slagging and fouling characteristics and lower ash fusion temperatures than coal. Units using a high slagging coal and having high rates of volumetric heat release are not considered good candidates for RDF cofiring. Furnace slagging and ash buildup appear to be a result of insufficient or excessive RDF furnace injection velocities. However, slagging and fouling problems can be reduced by lowering the RDF ash content by additional screening at the processing plant. Reduced boiler efficiency will occur from RDF cofiring due to increased flue gas exit temperature (furnace slagging), increased flue gas flow rate (resulting from higher excess air, lower fuel Btu value and higher fuel moisture), and increased heat losses (attributable to dry gas loss, unburned combustibles, and fuel moisture loss).

In the 1988 study *Guidelines for Cofiring Refuse-Derived Fuel in Electric Utility Boilers*, the Electric Power Research Institute reported that RDF cofiring reduces boiler efficiency by 1.9 to 4.2 lower

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<sup>21</sup> IGT report footnote No. 18 (p 65).

heating value of fuel), higher boiler ash loading, clinkering and bridging of dump grates (due to lowered ash softening temperatures and insufficient excess combustion air), plugging of sluice lines, plugging of clinker grinder, and plugging of ash dewatering bin screen. Reduced electrostatic precipitator (ESP) efficiency has been attributed to increases in fly ash electrical resistivity and flue gas flow rate. The net capacity of coal-fired units may decrease during RDF cofiring due to ash handling and gas flow limitations in the boiler, ESP and induced draft fans. Finally, a potential exists for an increase in boiler metal tube wastage. The maximum RDF cofiring rate recommended by steam generator manufacturers is typically 20 percent of the total fuel heat input. This limit was selected to ensure that hydrochloric acid (HCl) concentration in the flue gas resulting from RDF combustion is low enough to avoid increased corrosion or tube metal wastage in the boiler.

## Biogasification

Anaerobic digestion is the conversion of biodegradable organic matter (paper, yard, and garden wastes) into biogas, composed of methane and carbon dioxide, through the action of bacteria in an anaerobic environment. The biological conversion process is the result of anaerobic bacteria degrading complex organic material (carbohydrates, proteins, fats) into simpler compounds (acetic acid,  $\text{CO}_2$ ,  $\text{H}_2$ ). These simple compounds, which are soluble in water, are then available to another group of anaerobic bacteria, the methanogens, and are converted to  $\text{CH}_4$ ,  $\text{CO}_2$ , and water. The resulting gas composition is usually 55 to 65 percent  $\text{CH}_4$  with the remainder  $\text{CO}_2$  plus a few minor components.

Approximately half of the energy content of the MSW can be converted to fuel gas, with the remainder being combusted in a manner similar to mass burning. The advantages of anaerobic digestion over mass burning include the production of a valuable fuel gas, reduction of atmospheric emissions by as much as 50 percent, and recovery of recyclable materials. In addition, more energy is recovered per ton of MSW input through anaerobic digestion than through combustion. In terms of cost, anaerobic digestion displays capital costs roughly equivalent to mass burning. Operating costs, however, are greater for digestion, due mainly to process power requirements. Digestion requires that the majority of inorganic material be removed from the solid waste stream prior to bioconversion. Passage of material that cannot be fermented through the system increases the size and operating cost of the fermenter and the residue disposal system. In particular, glass, sand, and metal greatly accelerate erosion and abrasion of equipment and plastic may cause plugging of pipes and pumps as well as problems with the mixing system.

Biogasification begins with preprocessing of the solid waste to achieve size reduction and separation of inorganic material. Removal of recyclables also takes place at this stage. Size reduction is important because smaller particles aid in the rapid bioconversion of the waste. The separated organic material, often called refuse derive feed (not fuel), has a moisture content of 20 to 50 percent. This must be increased to a moisture content of 75 percent or greater. Additional water is added by recycling reactor effluent, which conserves water, heat, nutrients, and alkalinity. Liquid recycle also makes it possible to operate the digesters over a wide range of solids concentrations by varying the amount of recycled filtrate. The flow reactors used in the digestion phase can be of different types: completely mixed reactor (CMR), continuously stirred tank reactor (CSTR), occasionally stirred reactor (OSR), and nonmixed vertical flow reactor (NMVFR) also referred to as SOLCON (solids concentrating). The reactor is filled with the organic feed and flooded to a specified solids concentration. Mixing plays an important role in successful digestion; the degree of mixing must be sufficient to maintain a relatively homogenous slurry. The methanogenic bacteria necessary for producing  $\text{CH}_4$  are very sensitive to pH and are generally inhibited when pH drops below 6.6. If alkalinity formed by the decomposition of the feed is not adequate, then additional alkalinity must be added to the reactor, usually in the form of lime. However, recycle of filtrate will reduce the need for additional alkalinity. As the fuel gas is evolved, it may be cleaned by a variety of methods in order to upgrade the gas to pipeline quality or substitute natural gas (SNG). This requires

the removal of contaminants such as water, hydrogen sulfide, and CO<sub>2</sub>. Alternatively, the biogas can be used to power a combined cycle gas turbine or engine driven cogeneration system.

Undegraded solids residue, equal to about 40 percent of the original amount of MSW, will remain after digestion. Several options exist for the disposal of this solid residue, including landfilling, composting, combustion, and thermochemical gasification. Of these, combustion seems most attractive because the remaining material contains large portions of high energy components such as lignin and plastics, and requires very little processing before combustion (drying via a screwpress is sufficient). The dewatered filtercake should burn at least as cleanly as RDF.

A previous study<sup>22</sup> has indicated that anaerobic digestion with residue combustion and a combined cycle gas turbine may be 35 percent more efficient at generating electricity than mass burning, due to the higher efficiencies involved in burning gas compared to burning MSW. One simple way of achieving a combined cycle is to send the gas turbine's hot exhaust gases to the solid residue boiler.

Anaerobic digestion has been successfully applied to MSW in several test projects. RefCoM (Refuse Converted to Methane) liquid phase reactor was operated in Pompano Beach, FL from 1978 to 1985.<sup>23</sup> The RefCoM facility used a continuously mixed digester and accepted up to 1000 tons per day of MSW and sewage sludge. The facility utilized two digesters, each with about 45,000 ft<sup>3</sup> of capacity, operating in parallel. Fixed cover tanks, variable speed mixers, and a vacuum filter system were installed. A reliable preprocessing module was developed that was designed for optimizing digester feed. The biological process proved to be very successful. Because it operates in the liquid phase, RefCoM is characterized by simple materials handling with respect to digester contents and effluent. One of the primary goals at the facility was to test the effect of various biological process factors, including hydraulic retention time (HTR), temperature, mixing, particle size, pH and nutrient levels. The facility showed that a digester operated at higher solids concentration requires less reactor volume, thus reducing both capital and operating costs. In addition, operating with a fermentation temperature in the thermophilic range, as opposed to the mesophilic range, also significantly reduces the retention time necessary for good conversion. In general, HRT and temperature were the major factors cited in process conversion efficiency.

Two European systems are designed to provide a high quality compost in addition to producing biogas. The DRANCO (dry anaerobic composting) system is a hybrid completely mixed reactor operated in the solid phase. A 3 ton/day pilot plant is now in operation in Gent, Belgium. The facility normally accepts the product from an MSW composting facility, but has been successfully demonstrated with a number of other substances including RDF. In the basic DRANCO system, the end product is dewatered, dried, and marketed as compost. Other arrangements are possible, including combustion of part of the residue. The VALORGA system is a completely mixed reactor that includes injections of compressed biogas. VALORGA is oriented toward maximum gas production in a relatively short retention time, at the expense of conversion efficiency and CH<sub>4</sub> yield. This indicates an emphasis on the sale of compost for revenue. A commercial facility using the VALORGA digester began operation in Amiens, France in 1988, with a capacity of 300 ton/day.

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<sup>22</sup> R. Legrand, T.M. Masters, and G.W. Fallon, "A Systems Analysis of the Biological Gasification of MSW and an Assessment of Proven Technologies," *Energy From Biomass and Wastes 13* (New Orleans, LA, 13-17 February 1989).

<sup>23</sup> H.R. Isaacson, J. Pfeffer, P. Mooij, and J. Gesalbracht, "RefCoM—Technical Status, Economics and Markets," Paper presented at the *Energy From Biomass and Wastes XII Conference* (New Orleans, LA, 15-19 February 1988).

## Ethanol Production

The production of ethanol from MSW is based upon the presence of cellulosic materials as part of the waste stream. The primary cellulosic feedstock for production of ethanol from MSW is wastepaper, which comprises nearly half of the total solid waste stream. The technology for ethanol production from cellulosic materials is fundamentally different from the more common production of ethanol from starch crops such as corn. Ethanol production from starch crops is performed through fermentation.

In addition to starch crops and cellulosic materials, ethanol can also be produced from high productivity energy crops (HPEC) such as hardwoods and herbaceous crops. The attraction of ethanol from MSW is the "negative cost" associated with receipt of tipping fees. This is in contrast to the other two feedstock options, which involve significant purchase costs to obtain the feedstock. Ethanol production from MSW also benefits from lower energy requirements to collect and process the feedstock materials.

Several approaches exist for the conversion of cellulosic materials to ethanol, differing primarily in the method of hydrolysis and the fermentation system used. Hydrolysis of cellulosic materials can be accomplished with acids or cellulase enzymes. Acid based hydrolysis of cellulose is very rapid but faces several limitations, including difficulty in controlling the reaction and the presence of an acidic sugar, which inhibits yeast fermentation. Byproducts of the acid hydrolysis process include furfural, CO<sub>2</sub>, and electricity that can be produced from solid residual matter. Enzymatic processes are far less developed technologically than acid based processes. However, enzymatic processes hold the promise of being significantly less expensive than acid-based processes in the long run. Recent cost reductions have stemmed from minimizing end product inhibition of cellulase, improved cellulase enzymes and fermentative microorganisms, and improved systems for xylose fermentation. Byproducts of enzymatic processes include CO<sub>2</sub>, fusel oil, yeast, and electricity from solid residual matter. Acid hydrolysis processes are being investigated by the University of Arkansas, the University of Missouri, and the Tennessee Valley Authority. Enzymatic processes have been developed by the Natick Research Group of the University of Arkansas, and the University of Alabama, among others.

The current enzymatic technology for conversion of cellulosic biomass to ethanol has the potential for significant improvements in the areas of pretreatment, biologically mediated process steps, and product recovery. Biological process steps are the most costly, the least well developed, and exhibit the greatest potential for improvement. Currently, the cost of enzyme constrains the reaction time to values far above the limit imposed by substrate reactivity. Improvements are being investigated that would lower the effective enzyme cost, including increasing the reactivity of the substrate after pretreatment, improving enzyme production systems, improving enzyme activity, and recycling enzyme. Such systems have the potential to produce enzyme at little or no incremental cost; the cost of the process becomes equal to that of hydrolysis and fermentation only, and radical cost reductions can be anticipated. Reactor design for high productivity solids conversion is another area with great potential for lowering the cost of biological processing.

## Ethanol Markets

Ethanol production in the United States currently amounts to about 800 million gal/year.<sup>24</sup> Low-level blends of ethanol and gasoline are the primary market for ethanol consumption in the United States. In fact, nearly 7 percent of all gasoline sold in the United States now contains fermentation-derived ethanol. The current selling price of ethanol is in the range of \$1.15 to \$1.20/gal. However, the effective

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<sup>24</sup> L.R. Lynd, J.H. Cushman, R.J. Nichols, and C.E. Wyman, "Fuel Ethanol From Cellulosic Biomass," *Science* (15 March 1991), Vol 251, pp 1318-1323.

price to the blender is reduced by more than \$0.50/gal by tax incentives. Without these tax incentives, fuel ethanol would not be cost competitive with petroleum. Ethanol is also used as a feedstock in the production of ethyl tertiary-butyl ether (ETBE). In addition to its primary use as a motor fuel additive, ethanol is used to make a wide variety of chemicals, including ethyl acrylate, vinegar, various glycol ethers, ethylamines, and ethyl acetate. Major U.S. producers of ethanol include Archer Daniels Midland, Eastman Chemical, and Union Carbide.



## 5 ECONOMIC EVALUATIONS

An economic evaluation was performed to determine the capital and production costs for 400 ton/day MSW processing facilities based on either the dilute acid hydrolysis (TVA) or the enzymatic hydrolysis (UAH) processes. The capital costs include direct costs (equipment, installation and piping, etc.) and indirect costs (engineering, supervision, and contractor's fee, etc.). Production costs include items such as raw material, labor, and utilities. The estimated total capital investment for the 400-ton/day MSW processing facilities based on the TVA and UAH processes are \$51,138,720 and \$63,245,666, respectively.

The evaluation considers both processes scaled to a capacity of 400 ton/day of MSW input. The 400 ton/day capacity matches one of the facility sizes evaluated by the TVA, but is much larger than the largest facility evaluated by UAH (27 ton/day). Thus, a greater degree of uncertainty will exist in the capital cost estimates for the UAH process.

The equipment costs reported in the TVA feasibility study are generally in agreement with those of typical ethanol production plants using acid hydrolysis, except the utilities section. Researchers estimated the capital costs of a boiler/generator system to produce steam and electricity from burning solid residues from the hydrolysis section to be \$3.42 million, as compared to \$1.6 million reported by TVA. Process calculations showed that 1 MW of electricity is produced in this boiler/generator. Based on a total electricity requirement of 3.2 MW, 2.2 MW or  $17.42 \times 10^6$  kWh/year of electrical power must be purchased; there is no excess electricity for export.

The TVA study proposed a molecular sieve system for ethanol dehydration. This system requires a cyclic adsorption/regeneration operation and a stream of hot air for adsorbent regeneration. A conventional azeotropic distillation system was used in the current study. The  $\text{CO}_2$  produced in the fermentation section was not credited as byproduct revenue, because the equipment costs for a  $\text{CO}_2$  recovery system were not included in the plant capital costs.

Calculations were made to verify the reported steam requirement. The TVA feasibility study showed that 181 ton/day of steam were used to evaporate and then condense 539 ton/day of water to recover 7 ton/day of solids from stillage. The TVA report states that a six-effect evaporator/condenser is used in this step. It may be possible to further reduce steam consumption by first filtering and then thermally drying the filter cake.

The equipment costs of the 400 ton/day MSW plant based on the UAH enzymatic hydrolysis process were scaled up from the UAH feasibility study for a 27 ton/day plant. Equipment capacity ratios were calculated from a revised material balance for a 400 ton/day MSW processing facility. Engineering calculations were performed to verify the scaled up costs of major equipment.

The capital cost of an ethanol distillation system was estimated because it was not included in the cost estimate of the 27 ton/day plant. The feasibility study proposed a novel pervaporation system for ethanol dehydration. For comparison to the TVA process, the conventional azeotropic distillation system was used and costed for the current study. The purchased cost of a boiler/generator system using solid residues recovered from the hydrolysis section as the boiler fuel was estimated to be \$2.3 million, compared to a scaled up value of \$4.08 million. The  $\text{CO}_2$  was not credited as a by-product and the costs of  $\text{CO}_2$  recovery were not included in the plant capital costs.

Labor costs and utility requirements, such as cooling water and electricity, were not shown in the 27 ton/day UAH study. The cooling water consumption and labor costs were assumed to be the same as

those for the TVA study. The electrical power requirement for the 400 ton/day UAH plant was assumed as the electricity for the 400 ton/day TVA facility plus the additional electricity required for equipment, such as centrifuges, pumps for ultrafiltration and reverse osmosis, vibrating screens and conveyors, etc. The total required electricity was estimated as 4.2 MW. Calculations showed that 0.7 MW of electricity is generated by the boiler/generator system, creating a net electric load of 3.5 MW with an annual purchased electrical consumption of  $27.72 \times 10^6$  kWh.

The economic evaluation begins with an estimation of direct capital costs. This includes cost estimates for purchased equipment, installation, instrumentation and controls, piping, electrical, buildings, yard improvements, service facilities, and land. Purchased equipment costs are separated according to MSW classification, hydrolysis, fermentation, ethanol recovery, and utilities and miscellaneous costs. Indirect capital costs consist of engineering and supervision, construction expenses, contractor's fee, project contingency, and a fund for working capital. The sum of direct and indirect plant costs is the total capital investment for the facility.

One important difference exists between this analysis and the evaluations performed by TVA and UAH. This analysis assumes U.S. Government ownership of the facility and that the facility will be located at a military site. Thus, factors such as depreciation, taxes, and insurance were not considered. The analyses performed by TVA and UAH assume private funding sources that require a fixed rate of return and include tax payments, depreciation of fixed capital investments, and facility insurance. Federal life cycle costing procedures also dictate the adoption of a "constant dollar" analysis. This is an analysis that does not consider the effects of inflation on process economics. All dollar amounts in the analysis represent current costs that are discounted by a "real" interest rate which does not include an inflation component. This is realistic if all parameters in the analysis are influenced by inflation to the same extent. However, landfill costs are expected to greatly outpace inflation for the foreseeable future, which will have important ramifications for the economics of the two processes under consideration. This issue will be discussed at length in the section on sensitivity analysis.

The annual operating costs and revenues for the facilities are presented in three sections: raw materials, other annual costs, and revenue. Other annual costs include charges for electricity, landfilling of residual wastes, labor, maintenance, and plant overhead costs. Table 12 presents a compilation of the cost assumptions used in the economic analysis. These costs were verified by a current (1 September 1991) issue of the *Chemical Marketing Reporter*. However, only minor differences were observed between current market prices and the prices used in the analysis of the TVA process, so many of the TVA cost assumptions were replicated in this analysis. The issue of operating cost changes will be dealt with more extensively in the sensitivity analysis section.

The primary revenue components for both facilities are MSW tipping fees and the sale of ethanol. Among the recyclables, sale of aluminum is the largest revenue generator. Other significant revenues are derived from the sale of furfural for the TVA process and the sale of fusel oil for the UAH process. Solid fuel does not appear as a revenue item for the TVA or UAH processes because all solid fuel produced is needed to generate electricity for facility consumption. Revenue received from tipping fees and sale of ethanol appears to be much more dependable than the revenue from other sources, due to the uncertain nature of local market prices for recyclables and chemicals. One revenue item not included is the sale of carbon dioxide. Although carbon dioxide recovery was specified in the original plans of both processes, this analysis concluded that the revenue benefit from the sale of  $\text{CO}_2$  was greatly outweighed by the capital cost of the recovery equipment. Furthermore, ready markets for  $\text{CO}_2$  do not exist in all locations.

The economic analysis assumes that facilities will operate for a 20-year lifetime, at a rate of 24 hours/day and 330 days/year. Salvage value at the end of the facility life is assumed to be negligible. The analysis uses a discount rate of 4.6 percent, which is consistent with FY 92 U.S. DOE/DOD guide-

Table 12

## Economic Analysis Cost Assumptions

Cost Type	Amount
<b>Revenues/byproduct credits</b>	
MSW tipping fee (TVA & UAH)	\$45/ton
Ethanol (TVA & UAH)	\$1.15/gal
Yeast (TVA & UAH)	\$150/ton
Furfural (TVA)	\$800/ton
Acetic acid (TVA)	\$580/ton
Carbon dioxide (TVA & UAH)	\$10/ton
Fusel oil (UAH)	\$6/gal
<b>Recyclables</b>	
Aluminum (TVA & UAH)	\$1000/ton
Glass (TVA & UAH)	\$15/ton
HDPE (TVA & UAH)	\$140/ton
PET (TVA & UAH)	\$140/ton
Ferrous metals (TVA & UAH)	\$50/ton
<b>Operating costs</b>	
Sulfuric acid (TVA)	\$70/ton
Lime (TVA)	\$45/ton
Potas. dih. phosphate (TVA)	\$50/ton
Urea (TVA)	\$115/ton
Sodium sulfite (TVA)	\$475/ton
Denaturant (TVA)	\$0.50/gal
Enzyme (UAH)	\$1/lb
Ammonia (UAH)	\$115/ton
Electricity (TVA & UAH)	\$0.05/kWh
Process water (TVA & UAH)	\$500/million gal
Cooling water (TVA & UAH)	\$50/million gal

lines for analysis of federal energy management programs. The output of the analysis is presented in the economic input/output section. Three evaluation measures are presented: simple payback period, net present value, and internal rate of return. Simple payback period (SPB) provides an indication of the time required to recover the initial investment. SPB is given in years and yields the break even time for the project. However, SPB is naive in that it does not consider the time value of money (i.e., interest), and ignores all cash flows occurring beyond the payback year. Net present value (NPV) is a more accurate measure that computes the discounted value of all cash flows over the life of the project. Thus, NPV provides an indication of the net benefit or cost of the project in today's dollars. Any project with a positive net present value should be accepted; any negative NPV project should be rejected. Higher NPV's are preferable to lower NPV's. NPV is completely consistent with, and computationally equivalent to, the total life cycle cost (TLCC) measure which is used in U.S. government life cycle cost procedures. The NPV terminology is used here rather than TLCC to denote that a positive economic benefit, rather than an economic cost, is being incurred. The final economic performance measure is the internal rate of return (IRR), often referred to simply as the rate of return. IRR provides an indication of the rate at which the investment earns a return over its lifetime, and is quoted as a percentage. IRR is mathematically equivalent to NPV, except that it is solved for a different variable; the IRR is actually the discount rate equated to a zero net present value. The actual rate of return from a project should be compared with the cost of capital in deciding whether to accept or reject a project. A project that displays an IRR higher

than the cost of capital should be accepted, while a project with an IRR lower than the cost of capital should be rejected. A higher percentage return is always better than a lower return. Both NPV and IRR can be ranked for different projects to determine the relative attractiveness of investment alternatives.

### **Waste Reduction Considerations**

The TVA and UAH processing facilities represent possible approaches toward mitigating the major problem of solid waste disposal. However, it should be noted that both facilities do not eliminate waste; rather, they reduce the quantity of waste that ultimately requires disposal. This is achieved through conversion of paper products into ethanol, as well as transforming the waste into residual solid fuel for use in combustion. In gross terms, the TVA process is capable of reducing 132,000 tons of MSW/year into 44,507 tons that still require landfill, a reduction of 66 percent. The UAH process realizes an even larger gross reduction of 85 percent, from 132,000 tons to 19,800 tons annually. However, some of this reduction is attributable to the separation and sale of recyclables. Specifically, both processes are designed to sell 30,115 ton/year of aluminum, glass, ferrous metals, HDPE and PET recyclable materials. Evaluated on a "net of recyclables" basis, the TVA process achieves a reduction of 43 percent and the UAH process achieves 62 percent reduction. As the composition of MSW changes and larger portions of recyclable materials are removed from the waste stream, the TVA and UAH processes will achieve a waste reduction closer to the net amount than the gross amount. The net waste reduction of both processes can be compared to other solid waste reduction/minimization strategies to determine the effectiveness and cost efficiency of various strategies. A comparison between the TVA and UAH processes on a waste minimization basis shows that the UAH process reduces a greater quantity of waste, in the amount of 24,707 ton/year, and achieves a net reduction of 62 percent compared to 43 percent for the TVA process.

### **Dilute Acid Hydrolysis Process Economics (TVA)**

Direct capital costs for the TVA process were increased over original TVA estimates for the 400 ton/day plant. Revisions were made to recognize higher utility capital costs for the boiler and turbogenerator equipment (\$3,419,974 rather than \$1,601,947). The remainder of purchased equipment costs were not changed; this includes MSW classification, hydrolysis and neutralization, fermentation, distillation/dehydration (ethanol recovery), and miscellaneous costs. Direct capital costs were increased for items including installation, instrumentation and controls, piping, electrical, buildings, yard improvements, service facilities, and land. These changes were made to reflect more accurate estimates according to standard ratio factors for capital investment items based upon delivered equipment cost for solid fluid processing plants. The total effect of all changes to direct capital costs was an increase from \$21,620,031 to \$30,767,238. Indirect capital costs were also increased for items such as engineering and supervision, construction expenses, contractor's fee, contingency, and working capital. Once again, changes to indirect capital costs were based upon standard ratio factors. Total capital investment for the 400 ton/day TVA process facility, including both direct and indirect capital costs, increased from \$35,078,283 to \$51,138,720. The total capital costs for the 400 ton/day MSW plant based on the TVA dilute acid hydrolysis process are summarized in Table 13.

Annual operating costs were reviewed and some changes were made. All raw material quantities and costs in the original TVA analysis were judged to be acceptable. These raw material costs included sulfuric acid, lime, yeast, potassium dihydrogen phosphate, urea, sodium sulfite, and denaturant. Costs for these commodities were checked against current market prices in the *Chemical Marketing Reporter*, and only minor differences were observed. Processing water and cooling water requirements were also acceptable. Landfill requirements, however, were increased from 35,095 to 44,507 tons to reflect an increase in quantities of material landfilled (primarily gypsum). This resulted in an increased cost of

Table 13

**Total Capital Investment for 400 ton/day MSW Plant  
Based on the Dilute Acid Hydrolysis Process (TVA)**

<b>Categories</b>	<b>Investment</b>
<b>Direct capital costs</b>	
Purchased equipment:	
MSW classification	1,485,150
Hydrolysis	3,651,580
Fermentation	1,144,400
Ethanol recovery	592,110
Utilities and misc.	3,627,524
Purchased equipment installation	4,095,298
Instrumentation and controls	1,365,099
Piping	3,255,237
Electrical	1,050,076
Buildings	3,045,222
Yard improvements	1,050,076
Service facilities	5,775,420
Land	<u>630,046</u>
Total direct plant cost	30,767,238
<b>Indirect capital costs</b>	
Engineering and supervision	3,360,244
Construction expenses	3,570,260
Contractor's fee	1,890,138
Contingency	3,780,275
Working capital	<u>7,770,565</u>
Total indirect plant cost	<u>20,371,482</u>
Total capital investment	51,138,720

\$423,540/year, to a total of \$2,002,815. Landfilling costs for the TVA process represent the second largest annual operating cost (after labor), and have a significant impact on process economics. Labor, supplies, maintenance, and plant overhead were all accepted as given in the TVA analysis. It should be noted that labor costs (estimated to be \$2,136,039/year) are a significant operating cost and have an effect on process economics. This analysis was not able to conclusively verify the labor cost estimate, so additional analysis of this issue may be warranted. Plant maintenance and overhead (estimated at \$1,177,380 and \$1,068,019/year, respectively) are also significant contributors to annual operating cost. Annual operating costs and revenues and economic indicators are summarized in Table 14. Annual operating cost components are presented in the first pie of Figure 6. This shows that landfill and labor costs account for over half of the total annual operating costs of the TVA process facility.

Revenue components for the process were reviewed. MSW tipping fees represent 132,000 tons of solid waste per year accepted at the facility. Unlike the TVA analysis, the researchers' analysis assumed a fixed (current national average) tipping fee at \$45/ton. Thus, revenue from tipping fees accounts for \$5,940,000/year, which is the largest single revenue component of the process. Ethanol production of 1,860,834 gal/year was verified, which amounted to annual revenue of \$2,139,959 based on an ethanol price of \$1.15/gal. This is equivalent to an ethanol yield of 14.1 gal/ton of MSW processed. Production of furfural and acetic acid from the ethanol production process was verified, amounting to \$756,800 and \$256,940/year, respectively. Revenue from the sale of recyclable materials was also found to be accurate. Aluminum is the most important contributor to process economics among the recyclables, contributing \$1,663,000/year.

Table 14

**Operating Costs, Revenues and Economic Indicators for 400 ton/day MSW Plant  
Based on the Dilute Acid Hydrolysis Process (TVA)**

Categories	Quantity	\$/Unit	\$/Year
<b>Raw materials</b>			
Sulfuric acid (tons)	3,546	70	248,220
Lime (tons)	2,597	45	116,865
Yeast (tons)	165	150	24,750
Potas.dih.phosphate (tons)	83	50	4,150
Urea (tons)	331	115	38,065
Sodium sulfite (tons)	165	475	78,375
Denaturant (gal)	88,611	0.5	44,306
<b>Other annual costs</b>			
Processing water (MG)	16	500	8,000
Cooling water (MG)	246	50	12,300
Electricity (kWh)	17,420,000	0.05	871,000
Landfilling	44,507	45	2,002,815
Labor	---	---	2,136,039
Supplies	---	---	176,607
Maintenance	---	---	1,177,380
Plant overhead	---	---	<u>1,068,019</u>
<b>Total annual costs</b>			<b>8,006,891</b>
<b>Revenue</b>			
MSW - tipping fees	132,000	45	5,940,000
Ethanol (gal)	1,860,834	1.15	2,139,959
Carbon dioxide (tons)	0	10	0
Furfural (tons)	946	800	756,800
Acetic acid (tons)	443	580	256,940
Electricity (kWh)	0	0.05	0
Aluminum (tons)	1,663	1,000	1,663,000
Glass (tons)	12,196	15	182,940
HDPE (tons)	3,386	140	474,040
PET (tons)	1,756	140	245,840
Ferrous metals (tons)	11,114	50	<u>555,700</u>
<b>Total annual revenue</b>			<b>12,215,219</b>
<b>Total annual net cash flow</b>			<b>4,208,329</b>
<b>Economic Input/output</b>			
Discount rate, %	4.6		
Lifetime, years	20		
Simple payback period, years	12.15		
Net present value, \$	3,131,676		
Internal rate of return, %	5.3		

Major changes were made to other revenue components, however. Recovery of CO<sub>2</sub> was eliminated for two reasons: (1) capital equipment costs in the TVA analysis did not include the cost of a CO<sub>2</sub> recovery system, and (2) the amount of CO<sub>2</sub> recovered (5837 ton/year) does not yield a sufficient annual revenue (\$58,367 at \$10/ton of CO<sub>2</sub>) to warrant installation of the costly recovery system. Thus, process economics will be improved by eliminating CO<sub>2</sub> recovery unless a more profitable market is found, which is unlikely. The other revenue item changed was the production of electricity from residual solid fuel.

## 400 ton/day facility

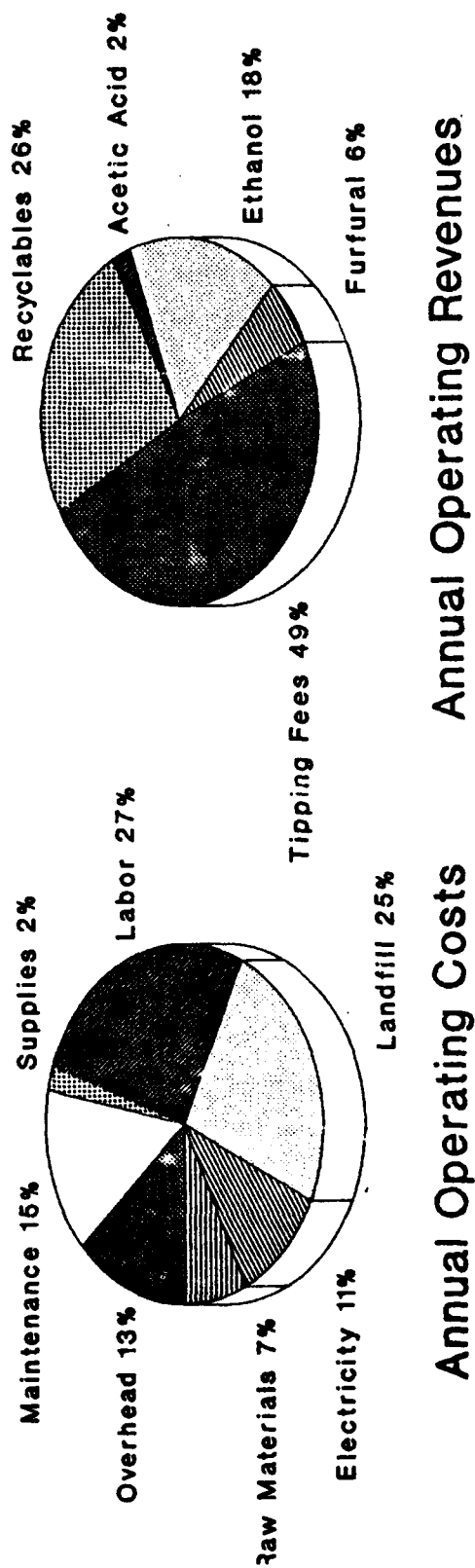


Figure 6. Annual Operating Costs and Revenues for a 400 ton/day MSW Plant Based on the Dilute Acid Hydrolysis Process (TVA).

The original TVA analysis indicated that combustion of residual solid fuel would meet all of the facility's process electric requirements of 3.2 MW, and would also provide  $1.2 \times 10^7$  kWh of electricity for sale. Thus, TVA assumed that annual operating costs of \$1,013,760 would be offset, and additional revenue of \$478,733 would be generated from the sale of electricity (priced at \$0.04/kWh). However, these electric estimates are in error. The researchers' analysis indicates that the facility will not be able to meet all of its own process electric requirements, and no additional electricity will be available for sale. Solid fuel combustion will account for only 993 kW of process electric requirements, forcing the facility to purchase  $1.74 \times 10^7$  kWh/year from an outside source. At a revised electric cost of \$0.05/kWh, this will cost \$871,000/year.

The second pie in Figure 6 shows the breakdown of annual operating revenues for the 400 ton/day TVA facility. Tipping fees account for half of total revenues, while sales of recyclable materials account for one quarter. Ethanol production accounts for only 18 percent of annual revenues. Thus, the conclusion can be drawn that this process is not really an "ethanol production facility" but rather an "MSW disposal facility." Income derived from refuse disposal services (e.g., tipping fees and recycling) generate 75 percent of facility revenues, while sales of byproduct chemicals (furfural, acetic acid, and ethanol) generate only 25 percent of revenues. Thus, tipping fees and recyclable material prices are much more important for facility revenue than are ethanol prices.

The economic evaluation of the TVA ethanol production process from MSW produces the following results: at a discount rate of 4.6 percent and a facility lifetime of 20 years, the process displays a simple payback period of 12.15 years, a net present value of \$3,131,676, and an internal rate of return of 5.3 percent.

#### Enzymatic Hydrolysis Process Economics (UAH)

Capital costs for the 400 ton/day facility based on the UAH enzymatic hydrolysis process were scaled up from the 27 ton/day plant size specified in the original UAH analysis. Due to the very large increase in plant size from what was originally intended, the capital cost estimates for the UAH facility are less certain than for the TVA facility. Purchased equipment costs are estimated to be \$12,986,790, including capital costs related to MSW classification, waste derived fuel (WDF) pretreatment and solids recovery (hydrolysis), fermentation, distillation/dehydration (ethanol recovery), utilities, and miscellaneous costs. Additional capital costs are required for installation, instrumentation and controls, piping, electrical, buildings, yard improvements, service facilities, and land. Total direct capital costs amount to \$38,051,293. Indirect capital costs account for \$25,194,373, which include engineering and supervision, construction expense, contractor's fee, contingency, and working capital. Total capital costs for the facility are estimated to be \$63,245,666. Thus, the capital investment required for the UAH enzymatic process is approximately \$12,000,000 greater than the capital investment for the TVA acid hydrolysis process. The total capital costs for the 400 ton/day MSW plant based on the UAH enzymatic hydrolysis process are summarized in Table 15.

Annual operating costs were also developed for the 400 ton/day enzymatic hydrolysis process. Enzyme, yeast, and ammonia are the raw materials required for operation of the UAH process. The cost of the required enzyme is not completely certain; it is estimated to cost \$5/lb at the present time but is expected to fall to \$1/lb. An enzyme cost of \$1/lb was used for this analysis. Other annual costs include processing water and cooling water, which were assumed to be equivalent to the values used in the TVA process. Labor, supplies, maintenance, and plant overhead were also set equal to the values established in the TVA analysis. Although cost estimates for these values were established using the acid hydrolysis process rather than the enzymatic process, it was decided that major operating cost parameters including labor, maintenance, and plant overhead should be very similar for the two processes. Because these



Table 15

**Total Capital Investment for 400 ton/day MSW Plant  
Based on the Enzymatic Hydrolysis Process (UAH)**

Category	Investment
<b>Direct capital costs</b>	
Purchased equipment:	
MSW classification	1,485,150
Hydrolysis	7,714,550
Fermentation	526,310
Ethanol recovery	900,000
Utilities and misc.	2,360,780
Purchased equipment installation	5,064,848
Instrumentation and controls	1,688,282
Piping	4,025,905
Electrical	1,298,679
Buildings	3,766,169
Yard improvements	1,298,679
Service facilities	7,142,734
Land	<u>779,207</u>
Total direct plant cost	38,051,293
<b>Indirect capital costs</b>	
Engineering and supervision	4,155,773
Construction expenses	4,415,509
Contractor's fee	2,337,622
Contingency	4,675,244
Working capital	<u>9,610,225</u>
Total indirect plant cost	25,194,373
<b>Total capital investment</b>	<b>63,245,666</b>

factors represent a large portion of total annual operating costs, a sensitivity analysis on this issue is warranted. Electric requirements were studied in detail for the process. The process is expected to require 4.2 MW of electricity; combustion of residual solid fuel will provide 700 kW of total electric requirements. The remaining  $2.77 \times 10^7$  kWh of electricity must be purchased from an outside source at an annual cost of \$1,386,000 (for an electric cost of \$0.05/kWh). Landfill requirements are 19,800 ton/year, with an annual cost of \$891,000 (for a landfill disposal fee of \$45/ton). It should be noted that annual operating costs for the UAH process are expected to be approximately \$800,000 less than annual operating costs for the TVA process. Annual operating costs and revenues and economic indicators are summarized in Table 16. The first pie in Figure 7 shows the breakdown of operating costs for the UAH process facility. The two most significant factors are labor and electricity, accounting for 29 and 19 percent of operating costs, respectively.

In terms of revenue for this process, tipping fees for 132,000 ton/year of MSW amount to \$5,940,000 (at \$45/ton of MSW). Ethanol production of 1,942,290 gal creates revenue of \$2,233,634/year. This is equivalent to an ethanol yield of 14.7 gal/ton of MSW. Sales of recyclable materials are exactly the same as in the TVA process, with aluminum representing the largest revenue contributor. Sale of 347 ton/year of fusel oil could generate up to \$611,067. However, the actual quantity of fusel oil that could be generated by the process as a byproduct is in doubt. Although fusel oil revenues were accepted for the base case, the issue will be revisited through a sensitivity analysis (Figures 8 and 9). Furthermore, the likelihood of finding local markets for fusel oil is not known. Recovery of CO<sub>2</sub> was specified in the original UAH process description, but was not considered in the researchers' analysis. Like the TVA

Table 16

**Operating Costs, Revenues and Economic Indicators for 400 ton/day MSW Plant  
Based on the Enzymatic Hydrolysis Process (UAH)**

	Quantity	\$/Unit	\$/Year
<b>Raw materials</b>			
Enzyme (ton)	155	2000	310,200
Yeast (ton)	46	150	6,930
Ammonia (ton)	914	115	105,122
<b>Other annual costs</b>			
Processing water (MG)	16	500	8,200
Cooling water (MG)	246	50	12,300
Electricity (kWh)	27,720,000	0.05	1,386,000
Landfilling (tons)	19,800	45	891,000
Labor	---	---	2,136,039
Supplies	---	---	176,607
Maintenance	---	---	1,177,380
Plant overhead	---	---	<u>1,068,019</u>
Total annual costs			<u>7,277,797</u>
<b>Revenue</b>			
MSW - tipping fees	132,000	45	5,940,000
Ethanol (gal)	1,942,290	1.15	2,233,634
Carbon dioxide (tons)	0	10	0
Fusel oil (tons)	347	1761	611,067
Yeast (tons)	1,082	150	162,300
Electricity (kWh)	0	0.05	0
Aluminum (tons)	1,663	1,000	1,663,000
Glass (tons)	12,196	15	182,940
HDPE (tons)	3,386	140	474,040
PET (tons)	1,756	140	245,840
Ferrous metals (tons)	11,114	50	<u>555,700</u>
Total annual revenue			<u>12,068,521</u>
<b>Annual net cash flow</b>			<b>4,790,724</b>
<b>Economic input/output</b>			
Discount rate, %	4.6		
Lifetime, years	20		
<b>Simple payback period, years</b>			
Net present value, \$	13.20		
	(1,464,728)		
Internal rate of return, %	4.3		

process, the large capital requirements of a CO<sub>2</sub> recovery system in the UAH process overwhelmed the minimal annual revenue generated by CO<sub>2</sub> sales. Revenue from sale of electricity was similarly disallowed. As mentioned in the UAH operating cost section, all residual solid fuel would be used to generate electricity to offset process requirements. No electricity would remain for resale. Total annual revenues for the UAH process, at \$12,068,521, are very close to the revenues for the TVA process.

## 400 ton/day facility

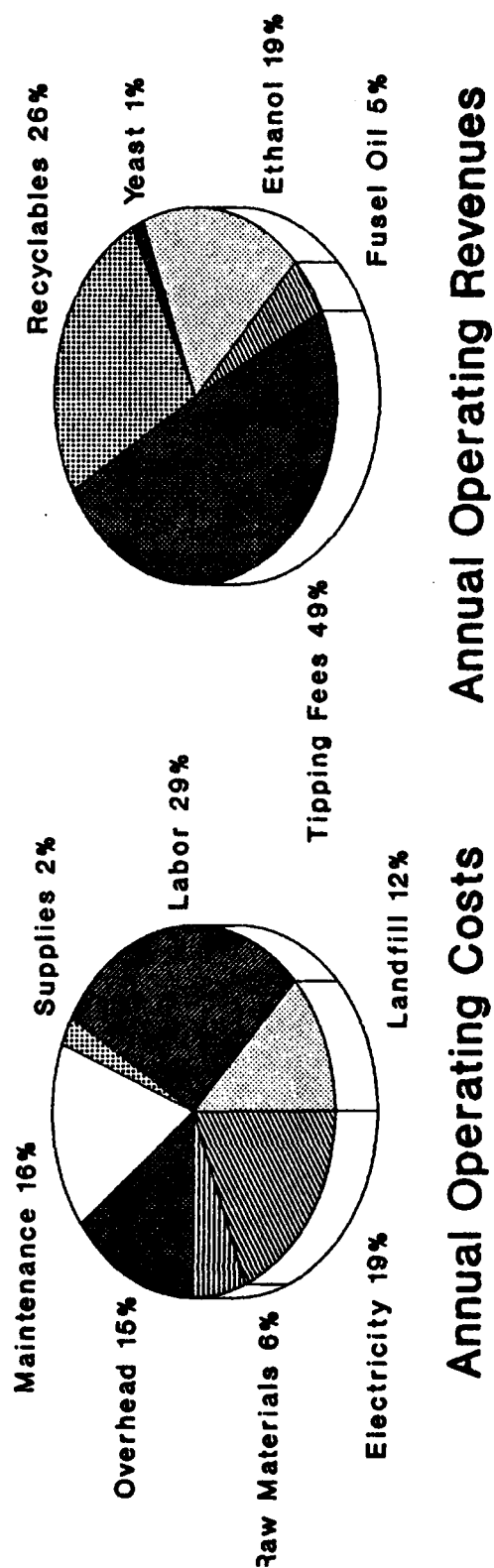


Figure 7. Annual Operating Costs and Revenues for a 400 ton/day MSW Plant Based on the Enzymatic Hydrolysis Process (UAH).

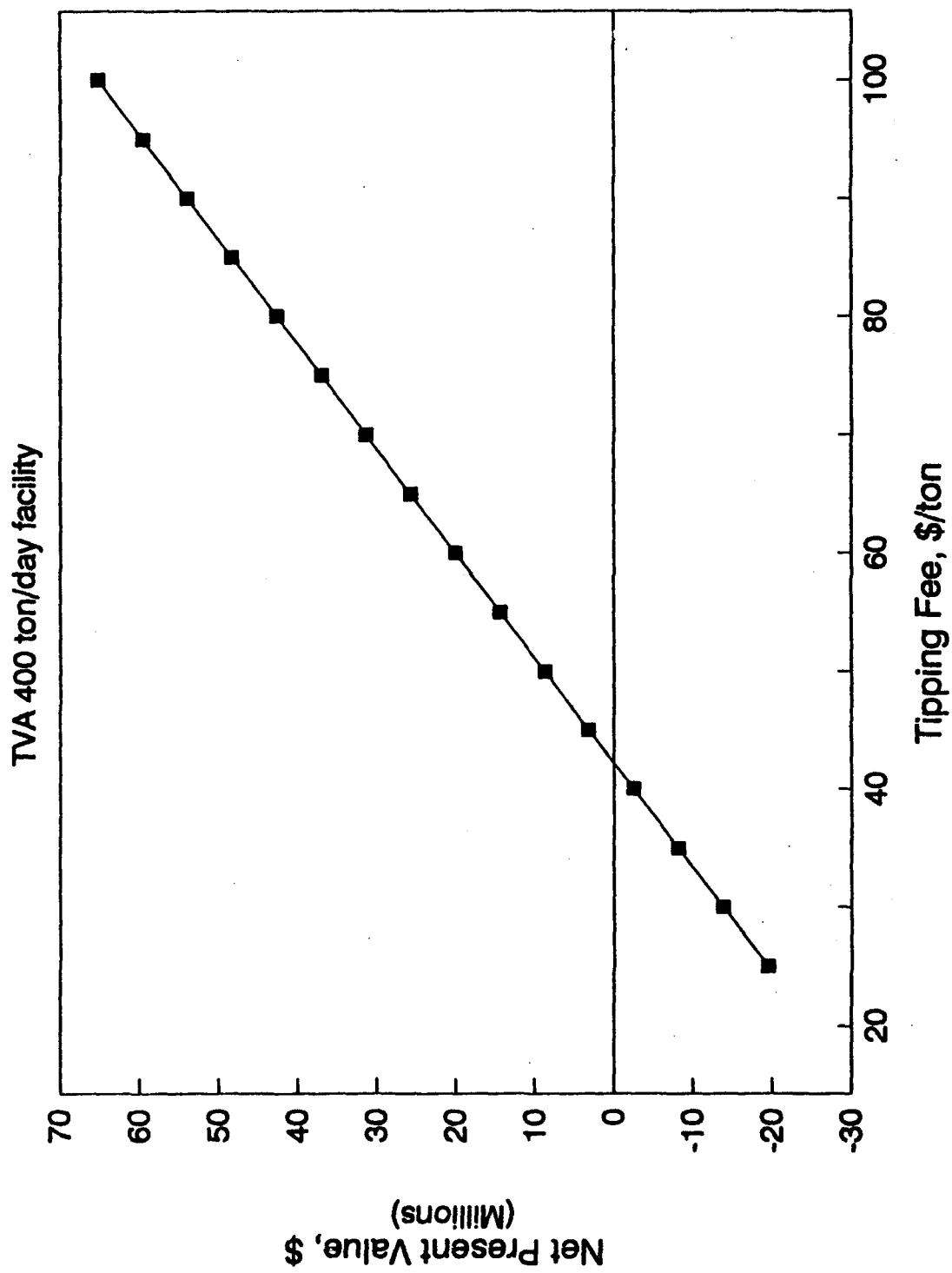


Figure 8. Sensitivity of Net Present Value to Tipping Fee (TVA).

Revenue components are shown in the second chart in Figure 7. Like the TVA process facility, the UAH facility derives 75 percent of its revenues from refuse disposal services (tipping fees and sale of recyclables) but only 25 percent from sales of byproducts (ethanol, fusel oil, yeast). Thus, tipping fees and recyclable material prices have a greater impact on facility economics than do ethanol prices.

The economic evaluation of the 400 ton/day MSW processing plant based on the UAH enzymatic process displayed the following results: a simple payback period of 13.2 years, a net present value of negative \$1,464,728, and an internal rate of return of 4.3 percent.

### **Sensitivity Analyses**

A sensitivity analysis of the TVA and UAH processes was conducted to isolate and highlight the effects of changes in key parameters. The sensitivity analysis will allow the reader to understand the effects that changes to one variable would have upon the base-case economic analysis. The analysis is also helpful in determining the areas where efficiency or cost improvements would yield the most beneficial effects on process economics.

#### ***Tipping Fee Sensitivity***

Tipping fees have a twofold effect on economics; they increase the revenues derived from MSW processing, and they increase costs due to landfilling residual waste matter that cannot be processed. In both the TVA and UAH processes, however, the former effect predominates. Both processes accept 132,000 tons of MSW for processing, and have remaining landfill requirements of 44,507 tons and 19,800 tons for TVA and UAH, respectively. Figures 13 and 14 show that process economics, as measured in net present value, improves considerably as the tipping fee is increased. In fact, the tipping fee is the most sensitive economic parameter for both the TVA and UAH processes. The TVA facility displays a breakeven tipping fee of approximately \$42/ton, while the UAH facility exhibits a \$47 breakeven point. The current national average tipping fee, and the one used in the base case analysis, was \$45/ton. The high degree of sensitivity is highlighted by the observation that raising the tipping fee by \$10/ton increases NPV by almost \$10 million for each facility. The UAH process showed a slightly higher sensitivity to landfill fees than the TVA process due to the greater net tipping fees received (tipping fee revenue minus landfill costs).

#### ***Tipping Fee Escalation Sensitivity***

This sensitivity analysis differs from the preceding sensitivity analysis in one important respect: the base case tipping fee remains at \$45/ton, but is escalated each year at a certain percentage rate. Thus, this analysis considers the case of a facility that experiences higher tipping fees with each year that passes. This assumption can be made because tipping fees are generally expected to rise much more quickly than inflation, as they have done in the recent past. Figures 10 and 11 show that process economics improve enormously if tipping fees increase consistently over time. The net present value of the TVA process would triple with a 5 percent annual increase in MSW tipping fees (above inflation). An annual tipping fee increase of less than 1 percent is sufficient to produce a breakeven point for the UAH processes, and an annual increase of 5 percent (above inflation) produces a NPV of nearly \$40 million. Very large net present values result from annual tipping fee increases in the range of 5 to 10 percent above inflation. This may be the most persuasive economic argument for the construction of MSW processing plants; annually increasing tipping fees should yield a steadily rising revenue stream over the life of the facility.

### *Ethanol Price Sensitivity*

The sensitivity of process economics to ethanol price is shown in Figures 12 and 13. The TVA process results in 1.86 million gal of ethanol/year, while the UAH process yields 1.94 million gal annually. As mentioned in the earlier discussion of operating revenues, ethanol production does not yield a very large fraction of the annual operating income for either process (18 percent for TVA and 19 percent for UAH) at the base case value of \$1.15/gal. Therefore, the lack of any great sensitivity to ethanol prices should not be surprising. Ethanol price breakeven points are observed at \$1.03 for the TVA process and \$1.22 for the UAH process. However, the y-axis scale shows that the change in net present value for a given change in ethanol prices is not very great. Thus, ethanol prices are not a major determinant in the economic feasibility of either process.

### *Electricity Price Sensitivity*

Electric prices are studied due to the heavy consumption of electricity by both processes, as well as the production of residual solid fuel for electric generation. The TVA process requires a net electric purchase of  $1.742 \times 10^7$  kWh, while the UAH process requires a net purchase of  $2.72 \times 10^7$  kWh. Thus, the UAH process is a larger consumer of electricity than the TVA process. Figures 14 and 15 present the sensitivity graphs for both processes, which show a medium to low degree of sensitivity. Breakeven points are observed at \$0.065/kWh for the TVA process and \$0.045/kWh for the UAH process. The negative slope of the electric sensitivity graph shows that both facilities are net purchasers of electricity, rather than net sellers as indicated in the original process descriptions. Reasons for this condition were presented previously. Electric price changes within a realistic range of values, however, are not sufficient to make a major change in the feasibility of either process. The current average electric price for industrial users is approximately \$0.05/kWh.

### *Aluminum Price Sensitivity*

The sensitivity of process economics to the price for recycled aluminum is presented in Figures 16 and 17. Aluminum prices were chosen since aluminum represents 53 percent of the revenue derived from sale of recyclable materials. Both processes realize 1663 ton/year of aluminum that is sold for recycle. Breakeven points are observed at approximately \$850/ton for the TVA process and \$1075/ton for the UAH process. The base case value for price of aluminum is \$1400/ton. Realistically, none of the recyclable material prices are capable of changing the economic feasibility of the process to a significant degree.

### *Annual Labor Cost Sensitivity*

Labor costs are examined because they represent a large and uncertain component of annual operating costs. Figures 18 and 19 show that annual labor costs have a significant impact on process economics. This is due to the recurring nature of the labor expenditure. Breakeven points are observed at \$2.35 million for the TVA process and \$2.0 million for the UAH process. These amounts are equivalent to 110 percent and 94 percent, respectively, of their current base-case levels. Process economics show a medium-high degree of sensitivity to annual labor costs. Labor cost reductions of \$200,000/year would translate into increases in net present value of approximately \$3.75 million for each facility. Increases in labor costs above the base case value of \$2,136,039/year are also considered in the sensitivity graphs.

### *Facility Lifetime Sensitivity*

Facility lifetime sensitivity, as illustrated in Figures 20 and 21, exhibits an upward sloping curve. This indicates that process economics improve for longer facility lifetimes. The base case lifetime value is 20 years for both facilities. Breakeven values are observed at 18 years for the TVA process and 21 years for the UAH process. The graph shows, however, that facility lifetime has a fairly small effect on facility economics in comparison to other factors in the analysis.

### *Landfill Quantity Sensitivity*

This sensitivity analysis examines the effect of increasing the amount of remaining solid waste that must be disposed of in a landfill after the processing steps are complete. This analysis assumes that the base case tipping fee of \$45 is paid for removal of remaining solid waste. The base case values for each process are 44,507 tons and 19,800 tons, respectively. Figures 22 and 23 show the effect that quantities of waste landfilled have upon facility economics. Breakeven points are observed at 50,000 ton/year for TVA and 17,000 ton/year for UAH. The sensitivity of the two processes to landfill quantities is approximately equal, although the TVA process appears to be more sensitive due to the much larger base-case landfill requirements.

### *Furfural Sensitivity — TVA*

The contribution of furfural to the economics of the TVA process is shown in Figure 24. For higher furfural production quantities, the net present value of the process improves. The base case assumes that 946 ton/year of furfural are produced, with a resale value of \$800/ton. The figure shows that a breakeven point is observed at 650 ton/year. However, the low degree of sensitivity evident in the graph indicates that furfural production is not a major determinant of process economics. A complete elimination of furfural revenue would decrease the net present value by approximately \$9.5 million, at which point the TVA facility would show a NPV of negative \$6.5 million.

### *Fusel Oil Sensitivity — UAH*

The contribution of fusel oil to the economics of the UAH process is shown in Figure 25. The base case assumes production of 347 ton/year of fusel oil, at a price of \$1761/ton. However, there is doubt that fusel oil would actually be a byproduct of the UAH process. The sensitivity graph shows the effect at a variety of fusel oil production levels. At the extreme of zero fusel oil production, net present value drops by almost \$7 million. No breakeven point was computed because increased fusel oil production above the base case value was not considered to be a realistic possibility.

### *Discount Rate Sensitivity*

The effect of changes in the discount rate used in the economic analysis is presented in Figures 26 and 27. The primary effect of the discount rate is to determine how heavily future cash flows are weighted in the analysis. Higher discount rates weight future cash flows less heavily than lower discount rates. The reader should remember that these are "real" discount rates which do not take into consideration the effect of inflation. The figures show that breakeven discount rates are observed at 5.25 percent for the TVA process and 4.5 percent for the UAH process. A medium-high degree of sensitivity is observed to changes in the discount rate. The base case discount rate of 4.6 percent was specified by the U.S. Department of Commerce, National Bureau of Standards for fiscal year 1992, which begins 1 October 1991. Real discount rates typical of private firms, in the range of 6 to 8 percent at the present time, produce negative net present values for both processes.

## Economic Conclusions

Neither process under consideration is primarily an ethanol production facility. Both are, in actuality, MSW disposal facilities because they derive 75 percent of their revenue from tipping fees and sale of recyclable materials. Ethanol production accounts for only 18 to 19 percent of total facility revenues. In terms of waste reduction/minimization, the UAH facility is somewhat more efficient than the TVA facility, with a net reduction of 62 percent rather than 43 percent. As a result of the heavy reliance on tipping fees for income, both processes display a very high sensitivity to tipping fees and escalation of future tipping fees. Breakeven tipping fees without escalation were observed for TVA and UAH of \$42 and \$47, respectively. However, annual escalation rates above zero in real terms are sufficient to make each process economically self sufficient. Prices for byproducts, including ethanol, were determined to be fairly insignificant to the overall feasibility of either process.

Annual expenditures for operating labor and maintenance should be reviewed for each process. These requirements are considerable, amounting to an estimated total of \$3.3 million for each facility/year. This represents 42 percent of total operating costs for the TVA facility and 45 percent for the UAH facility. These operating labor and maintenance estimates were based on inadequate information, resulting from the lack of existing facilities to use for comparison. More refined estimates for these two operating parameters would greatly increase the accuracy of the economic analysis.

The analysis also determined that capital costs for the facilities were large in relation to the estimated net annual cash flow (e.g., revenues minus costs). Capital costs would need to be reduced by \$1.5 million to make the UAH process economical under base-case conditions. The largest capital expenditures for both processes were in the areas of hydrolysis, utilities (power generation), and MSW classification. Capital expenditures for fermentation and ethanol recovery equipment were a relatively minor part of total expenditures.

In terms of the difference between the processes, the TVA process experiences significantly larger landfill requirements, while the UAH process experiences much greater expenditures for electricity. Neither facility benefits from the sale of electricity as originally projected, although both do offset a portion of electric requirements by self-generation with residual solid fuel. Total capital expenditures are approximately \$12 million greater for the UAH process than for the TVA process. Operating costs are roughly \$800,000 greater for the TVA process, while operating revenues are about \$200,000 higher for TVA. Total annual net cash flow is higher by about \$600,000 for the UAH process. The UAH process results in a longer payback by almost a year, a lower net present value by \$4.5 million, and a rate of return that is lower by 1.0 percent. Thus, the TVA process is marginally preferable. However, both cases are marginal in light of the degree of accuracy that can be expected from an economic analysis for commercial scale plants based only on pilot scale experimental experience. Under base-case conditions, the TVA process is acceptable because it realizes a small positive net present value, while the UAH process should be rejected due to a small negative net present value. The anticipation of escalating tipping fees would be the primary economic justification for constructing these facilities. Even small annual increases in tipping fees (in real terms) would be sufficient to make either facility experience a large positive net present value. Thus, one of the primary decision variables for this analysis is the anticipated escalation of tipping fees.



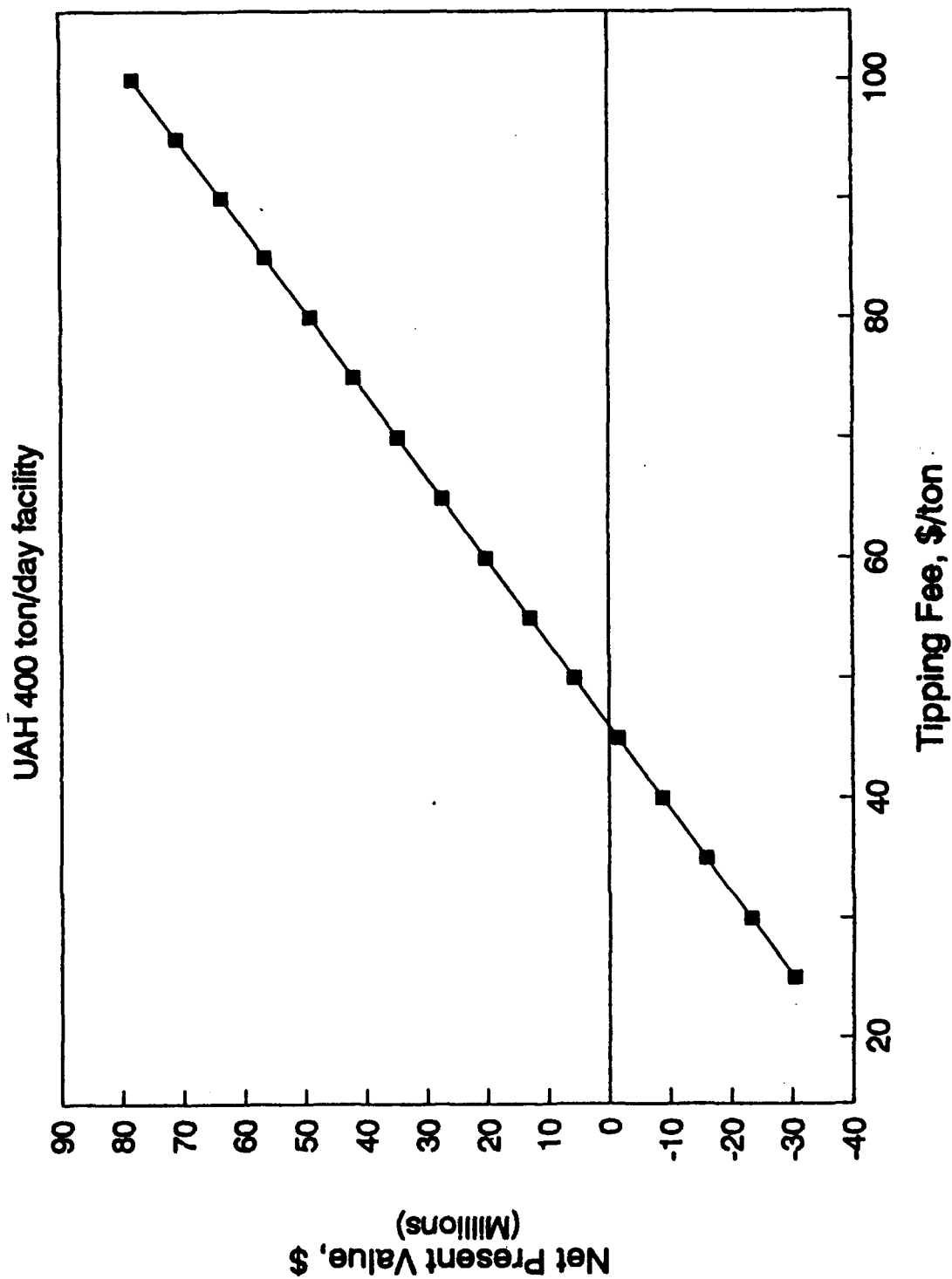


Figure 9. Sensitivity of Net Present Value to Tipping Fee (UAH).

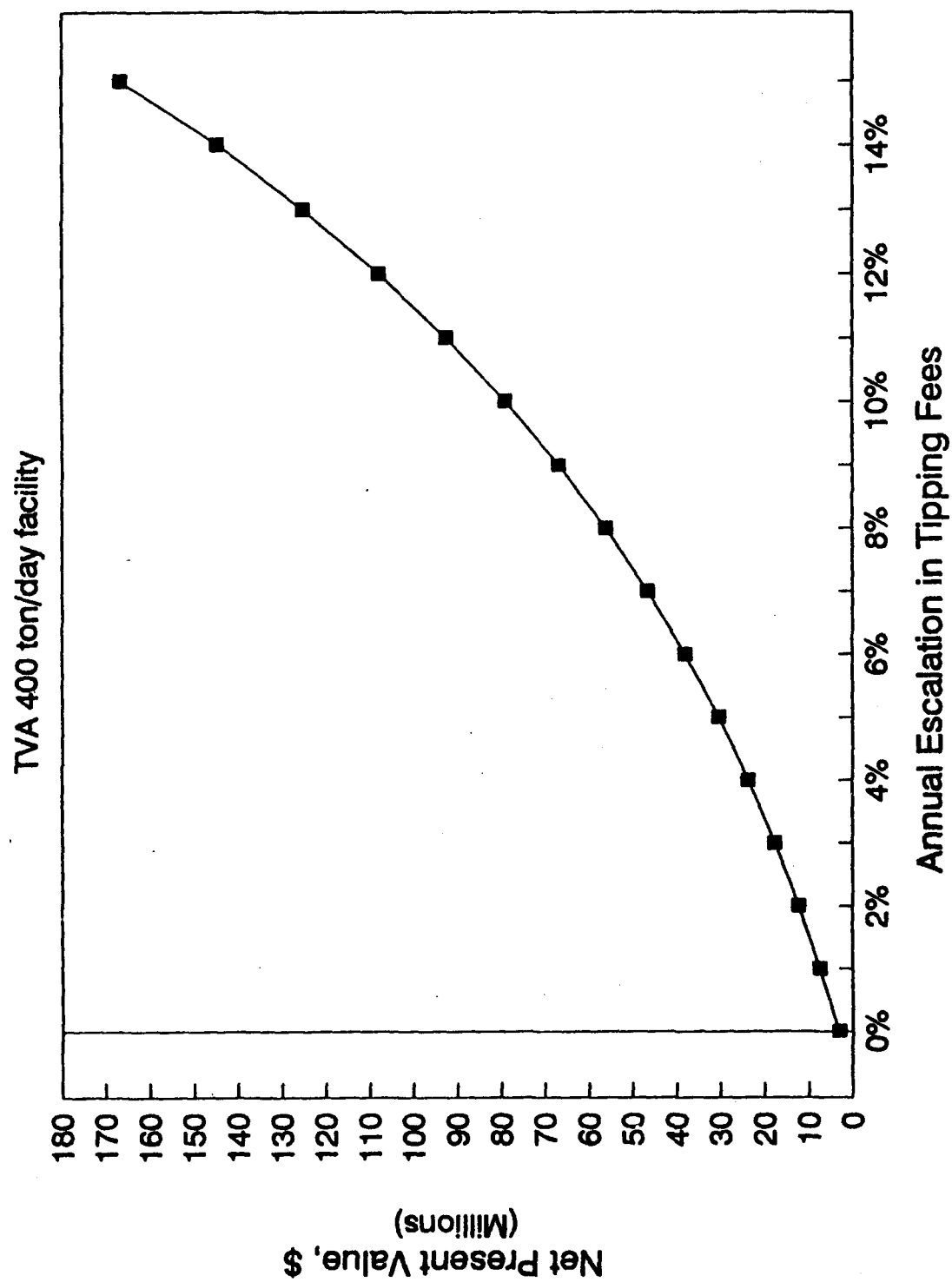


Figure 10. Sensitivity of Net Present Value to Tipping Fee Escalation (TVA).

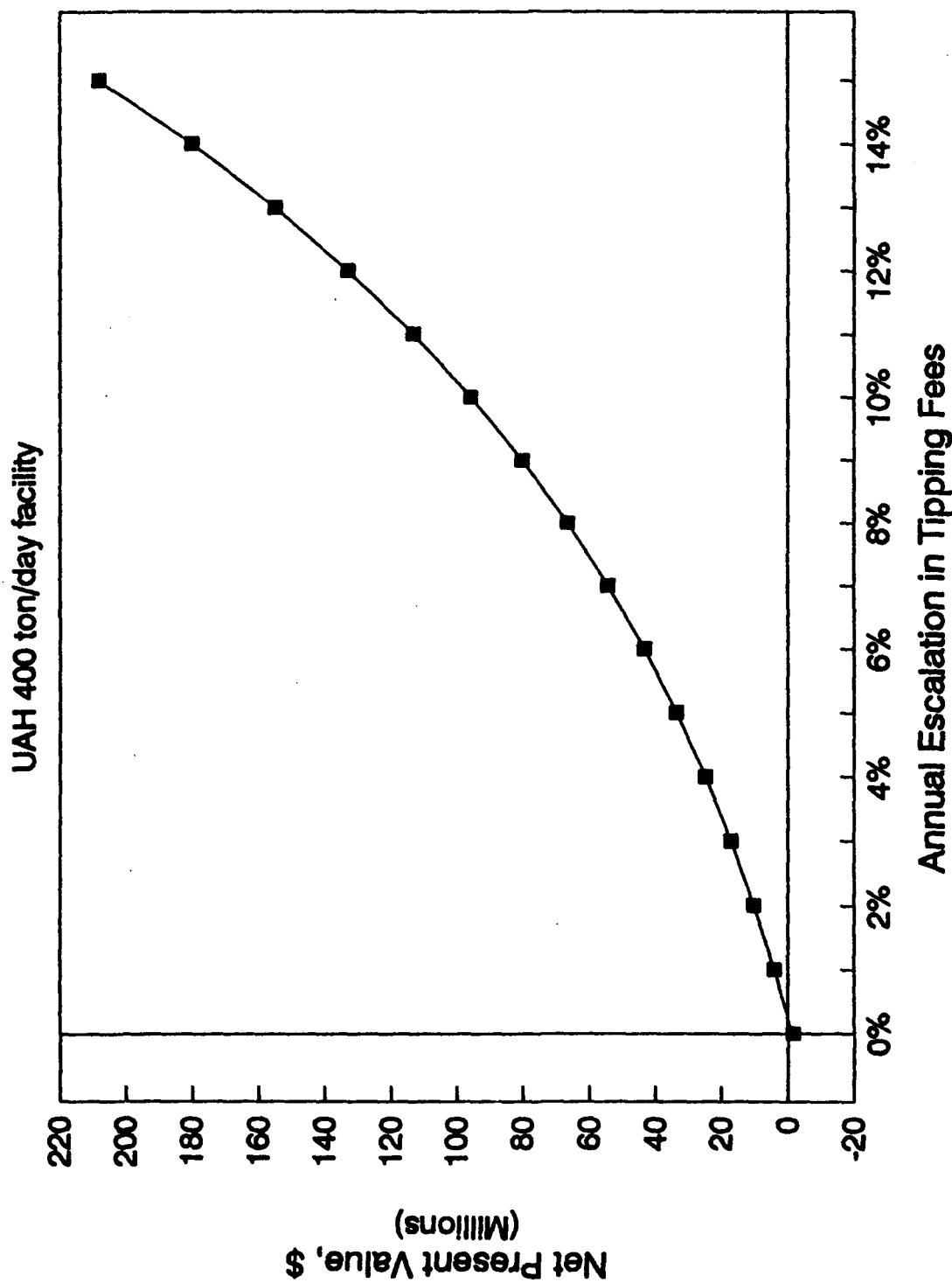


Figure 11. Sensitivity of Net Present Value to Tipping Fee Escalation (UAH).

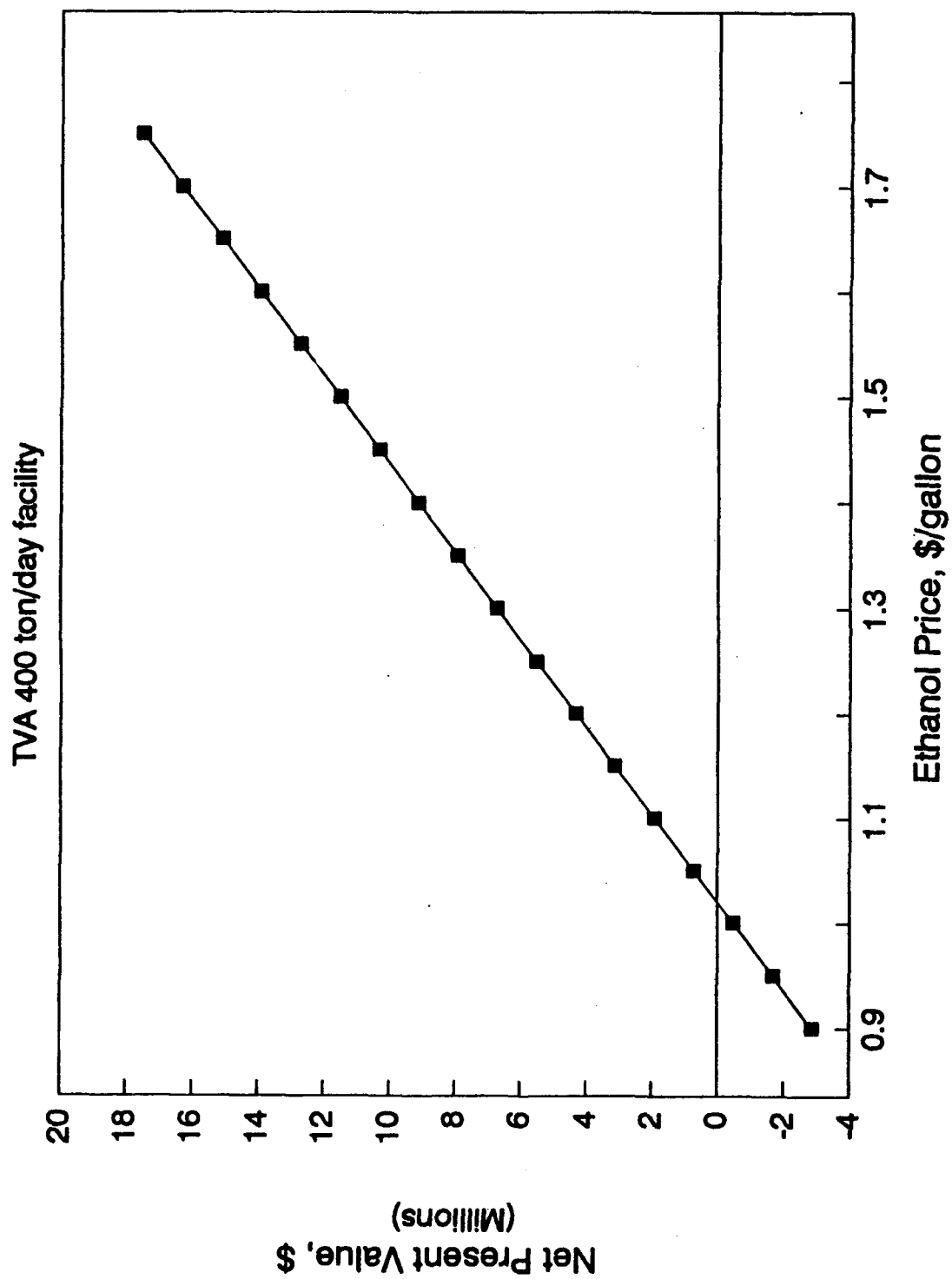


Figure 12. Sensitivity of Net Present Value to Ethanol Price (TVA).

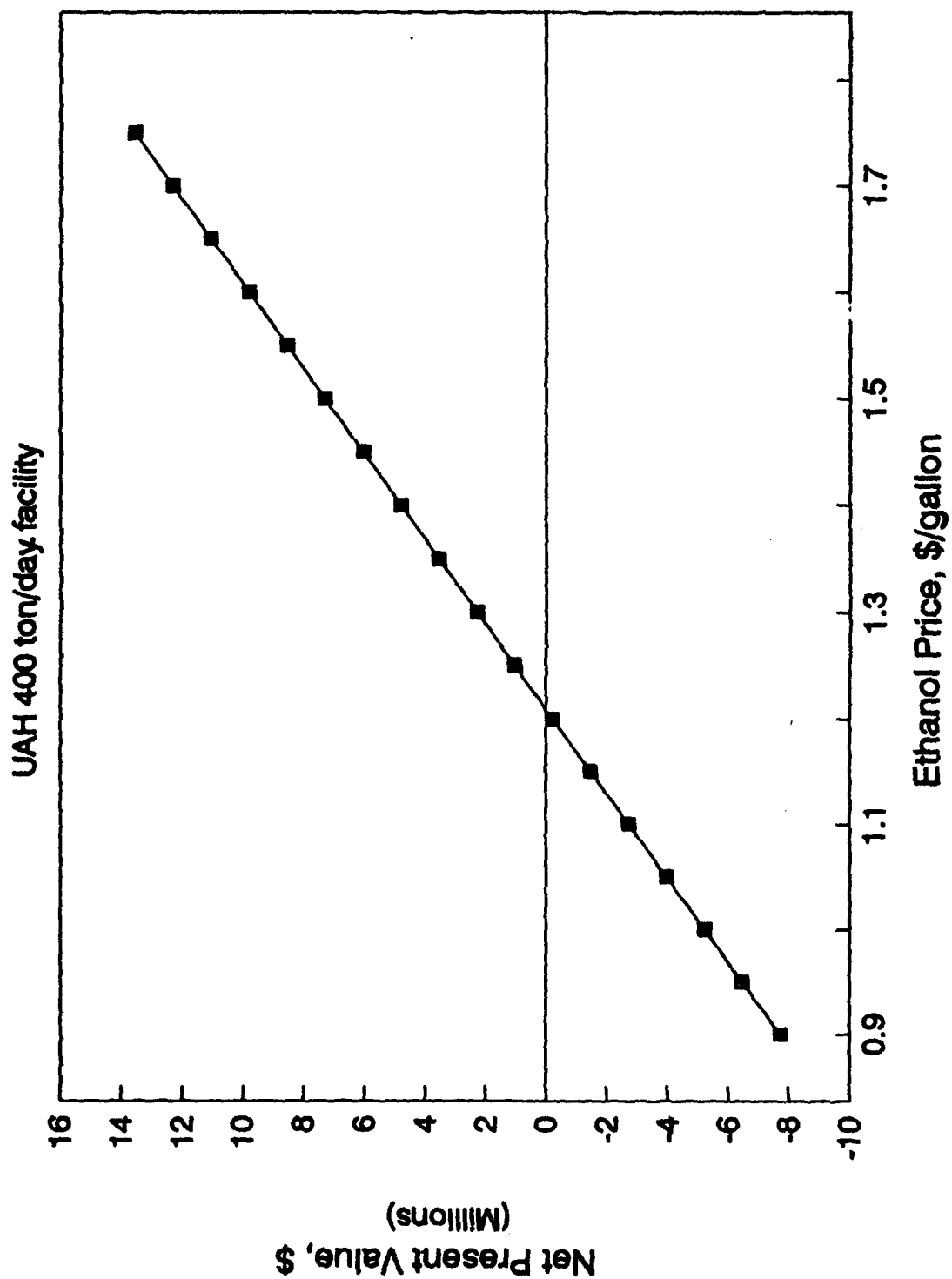


Figure 13. Sensitivity of Net Present Value to Ethanol Price (UAH).

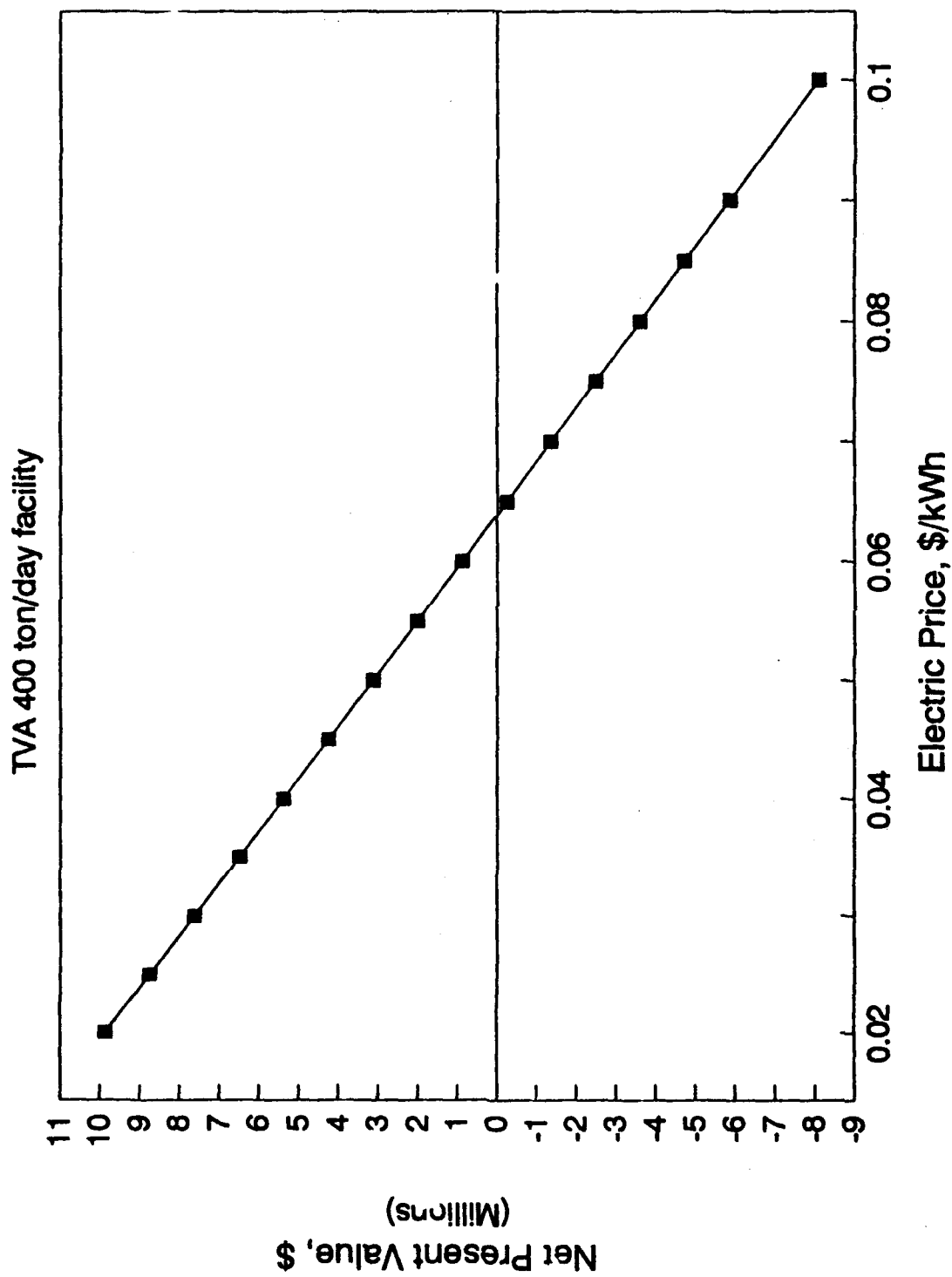


Figure 14. Sensitivity of Net Present Value to Electric Price (TVA).

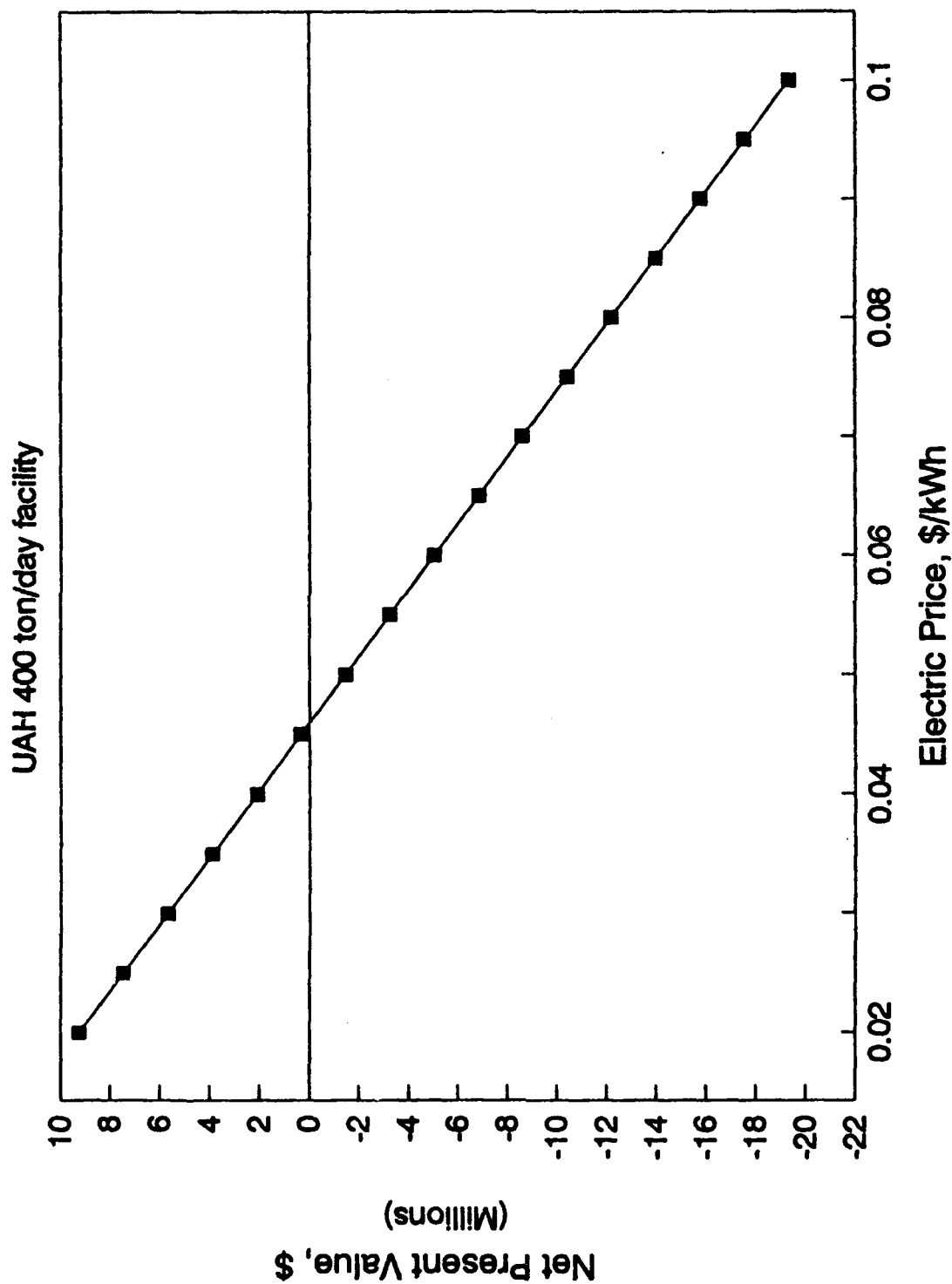


Figure 15. Sensitivity of Net Present Value to Electric Price (UAH).

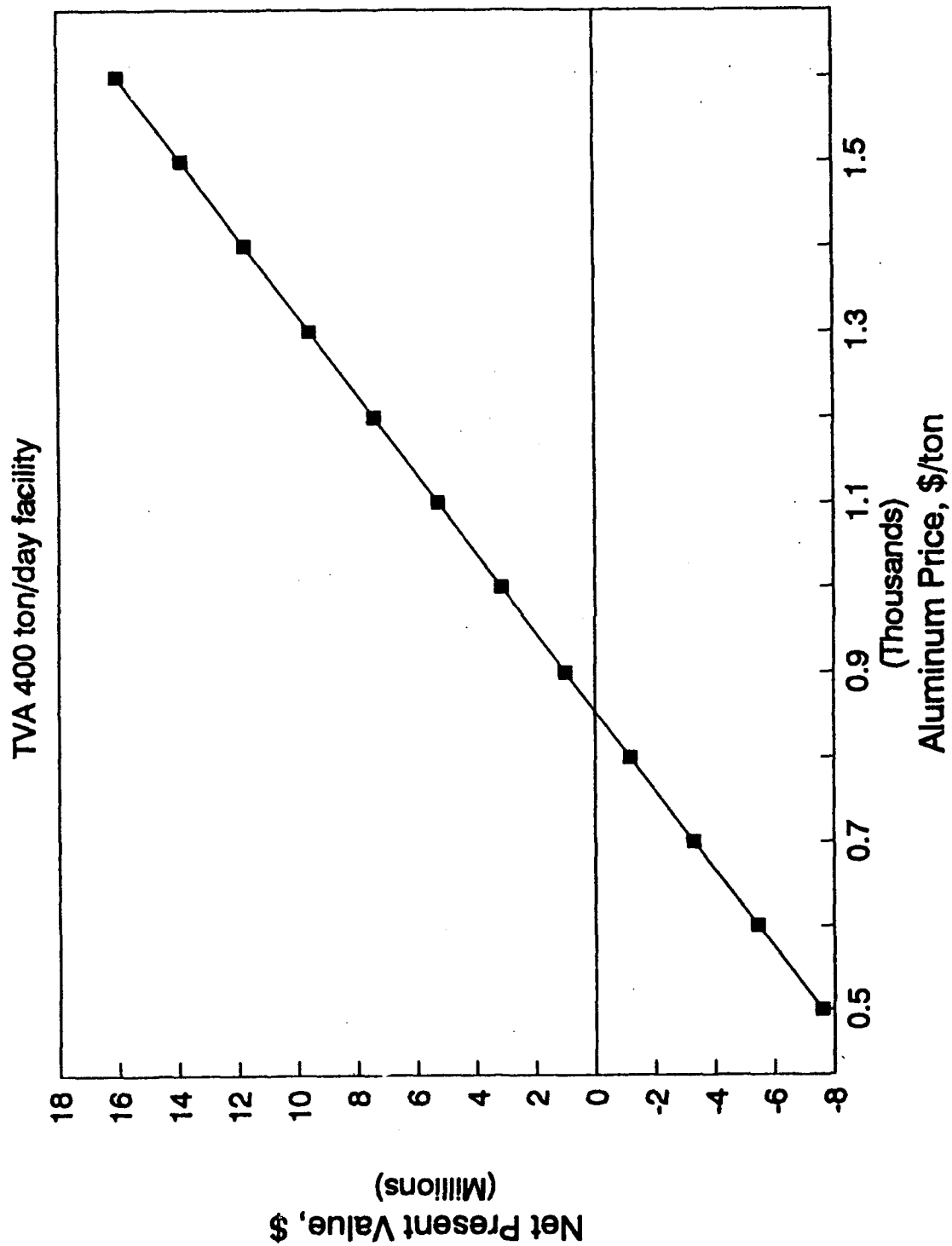


Figure 16. Sensitivity of Net Present Value to Aluminum Price (TVA).



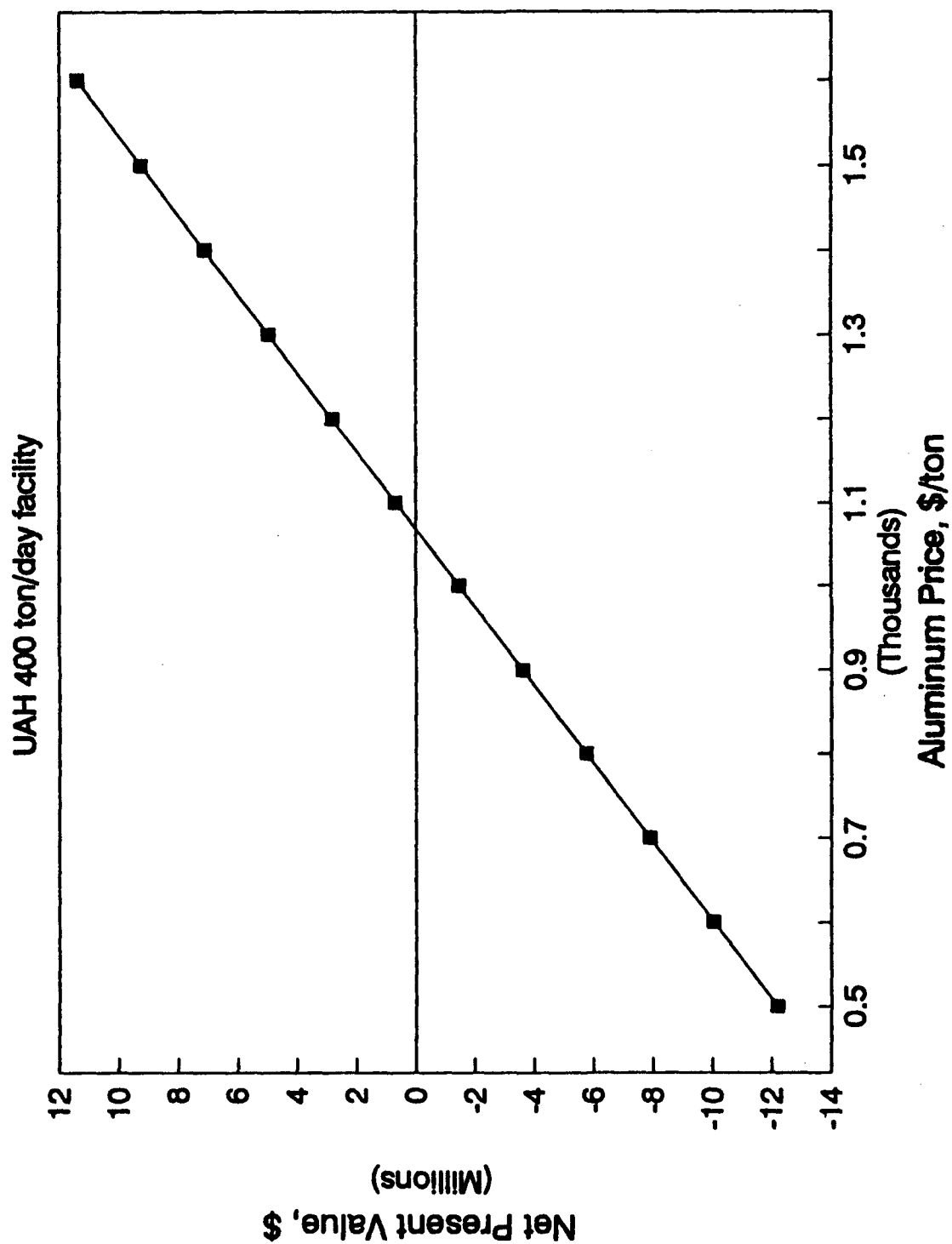


Figure 17. Sensitivity of Net Present Value to Aluminum Price (UAH).

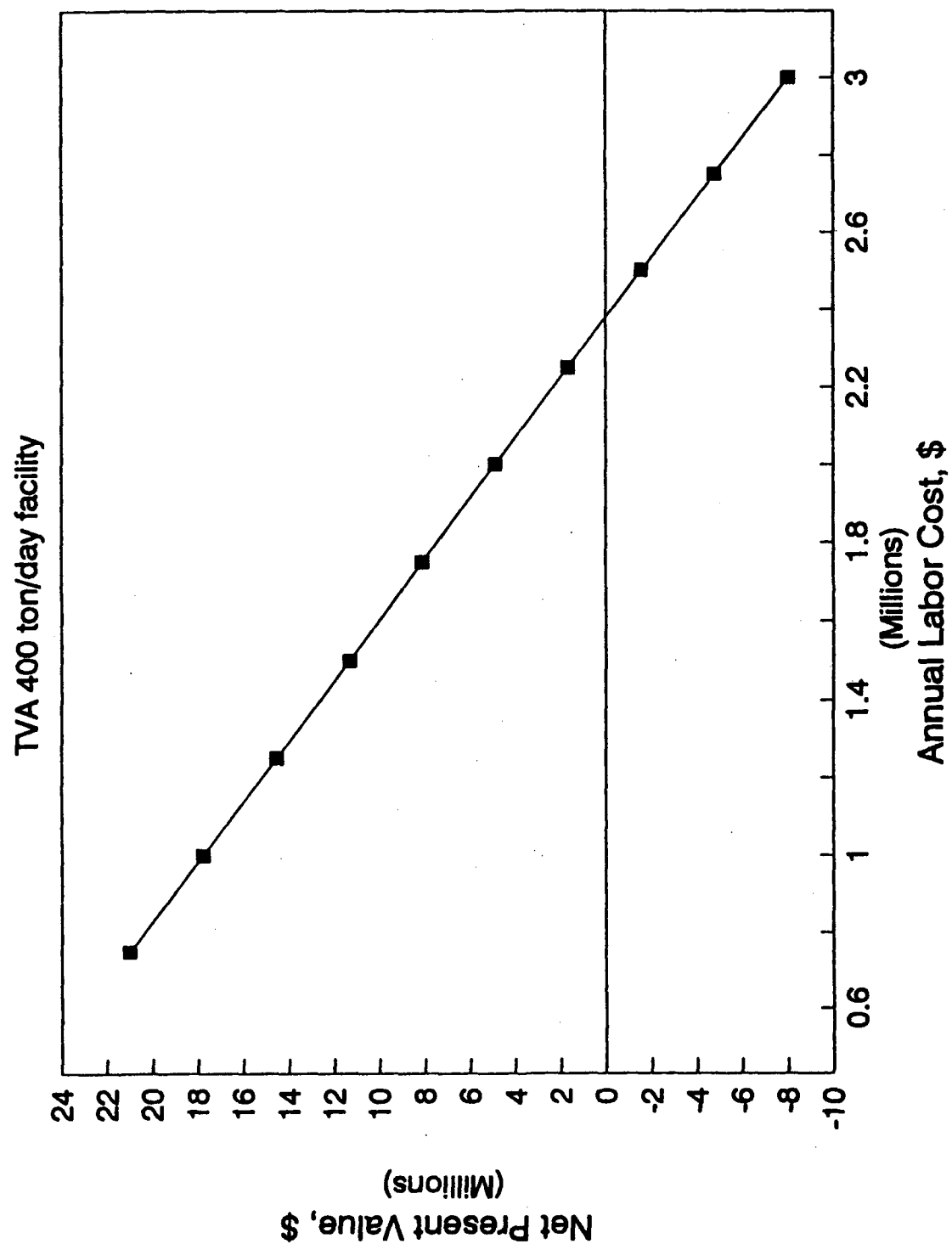


Figure 18. Sensitivity of Net Present Value to Labor Cost (TVA).

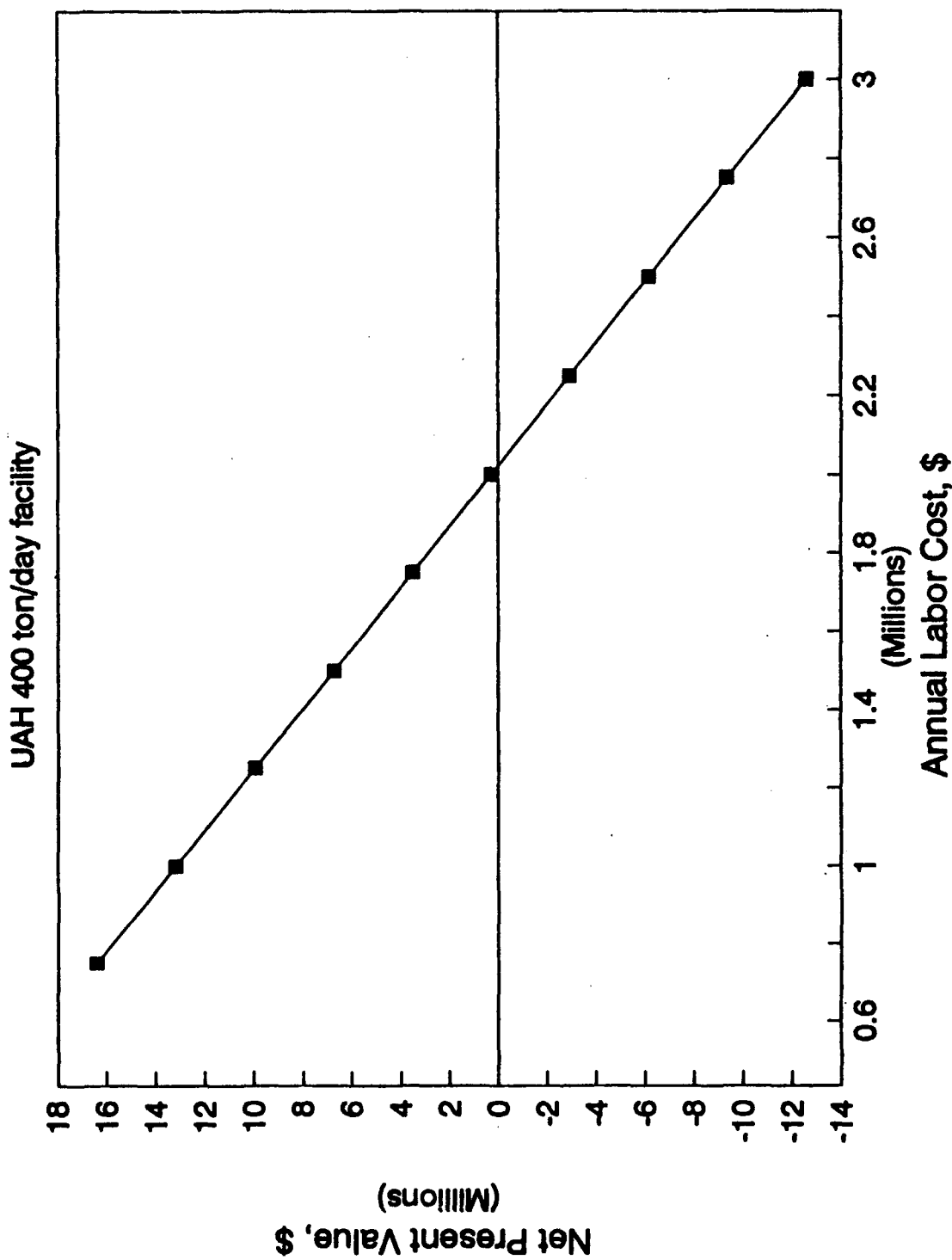


Figure 19. Sensitivity of Net Present Value to Labor Cost (UAH).

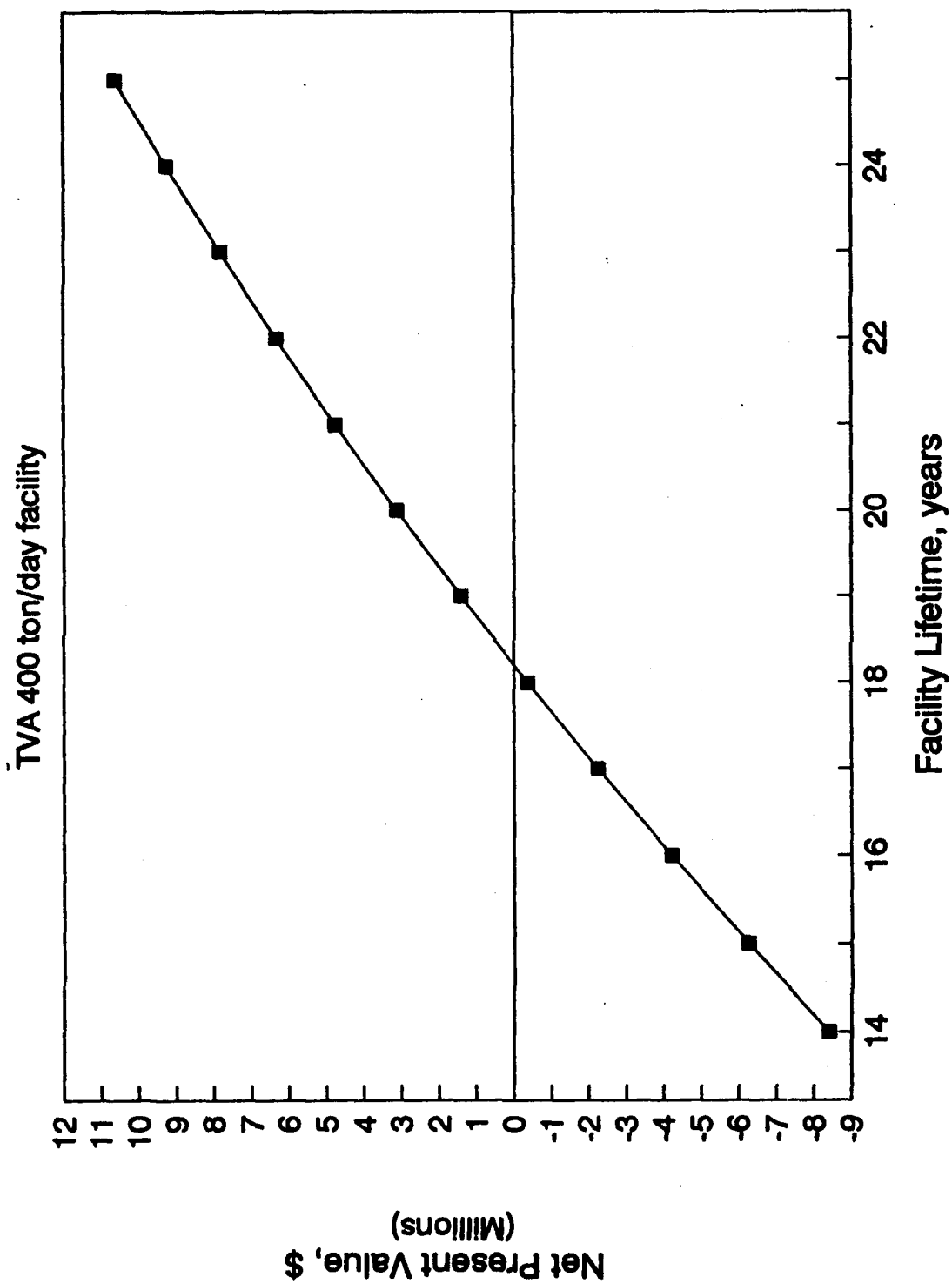


Figure 20. Sensitivity of Net Present Value to Facility Lifetime (TVA).

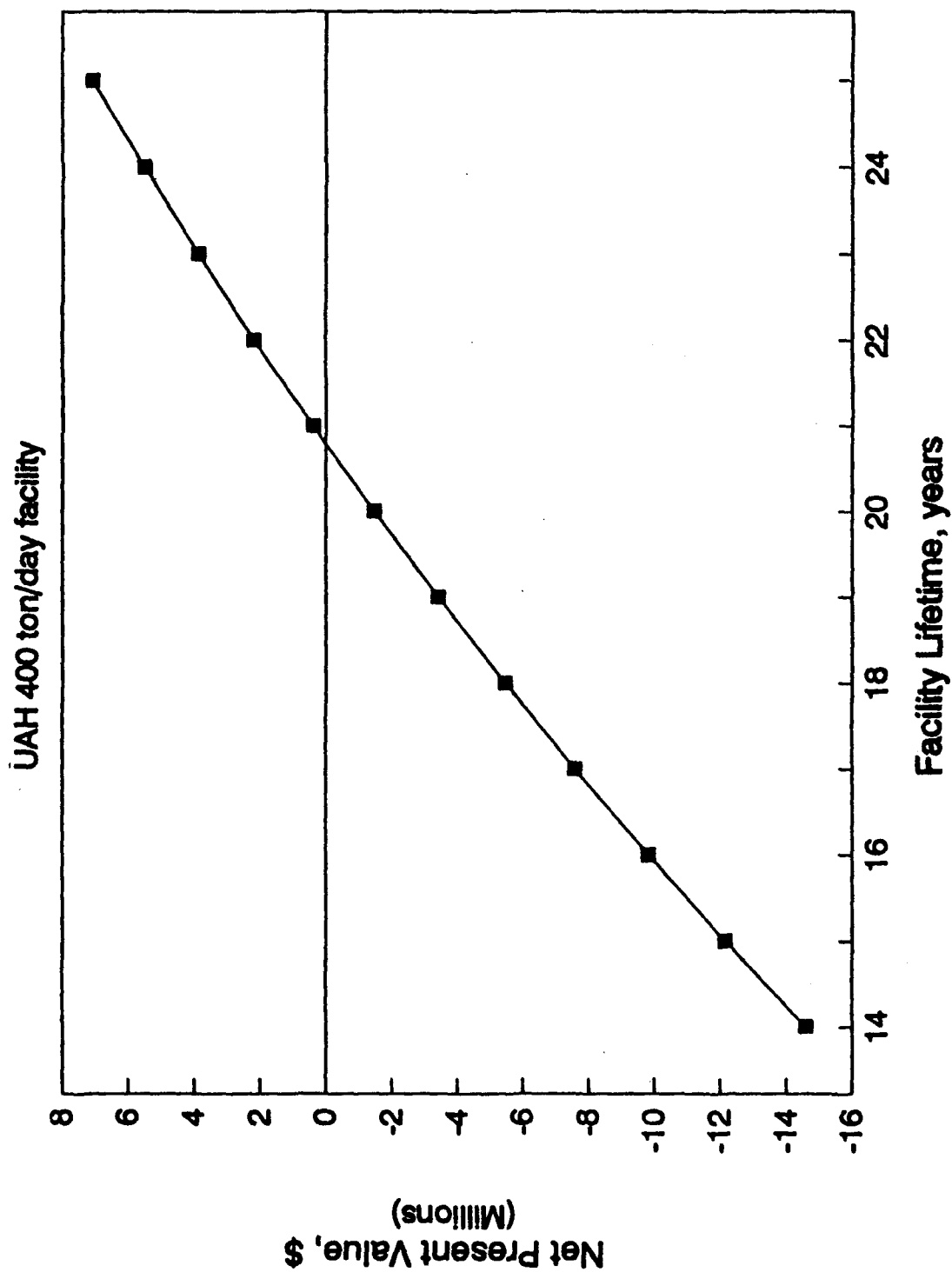


Figure 21. Sensitivity of Net Present Value to Facility Lifetime (UAH).

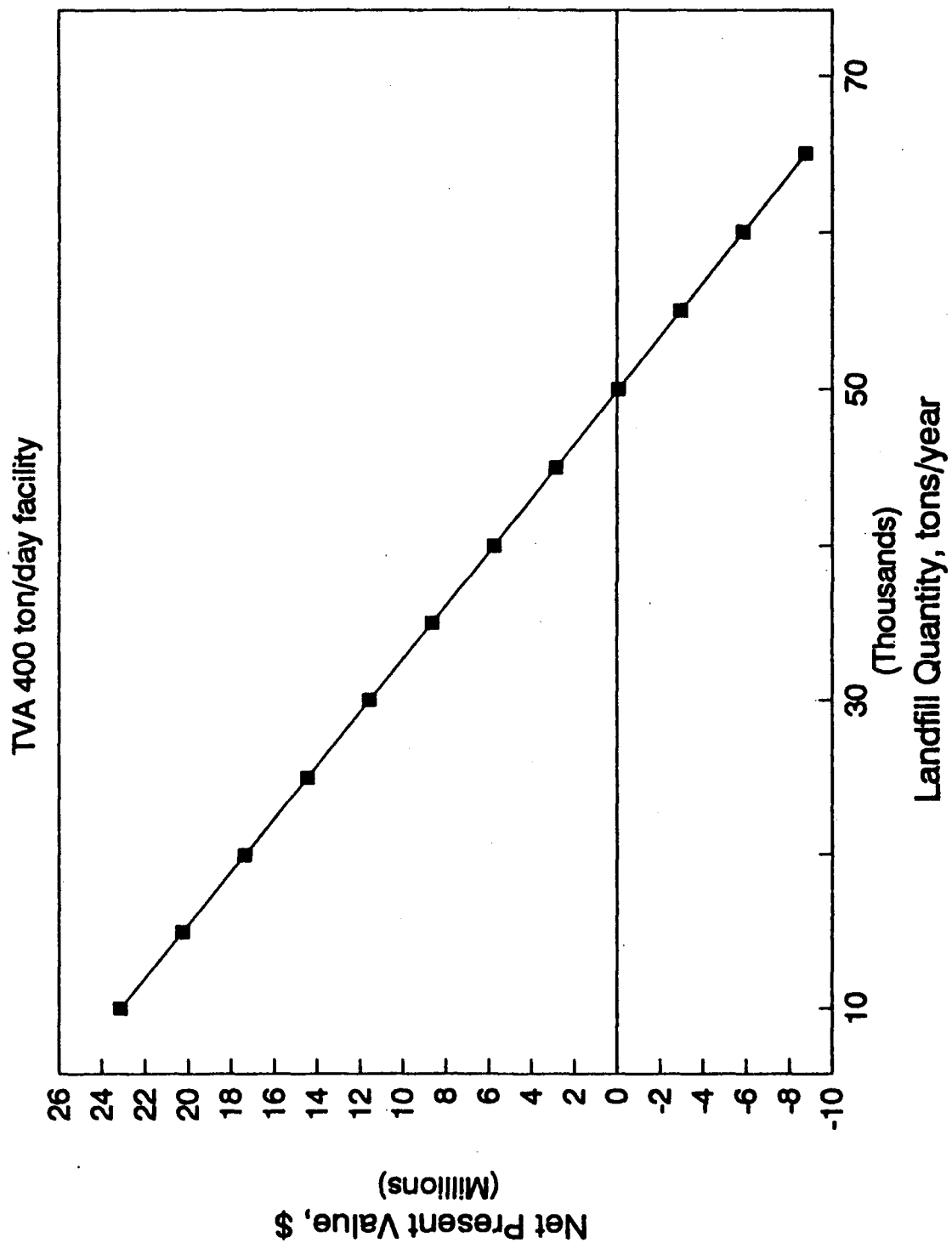


Figure 22. Sensitivity of Net Present Value to Landfill Quantity (TVA).

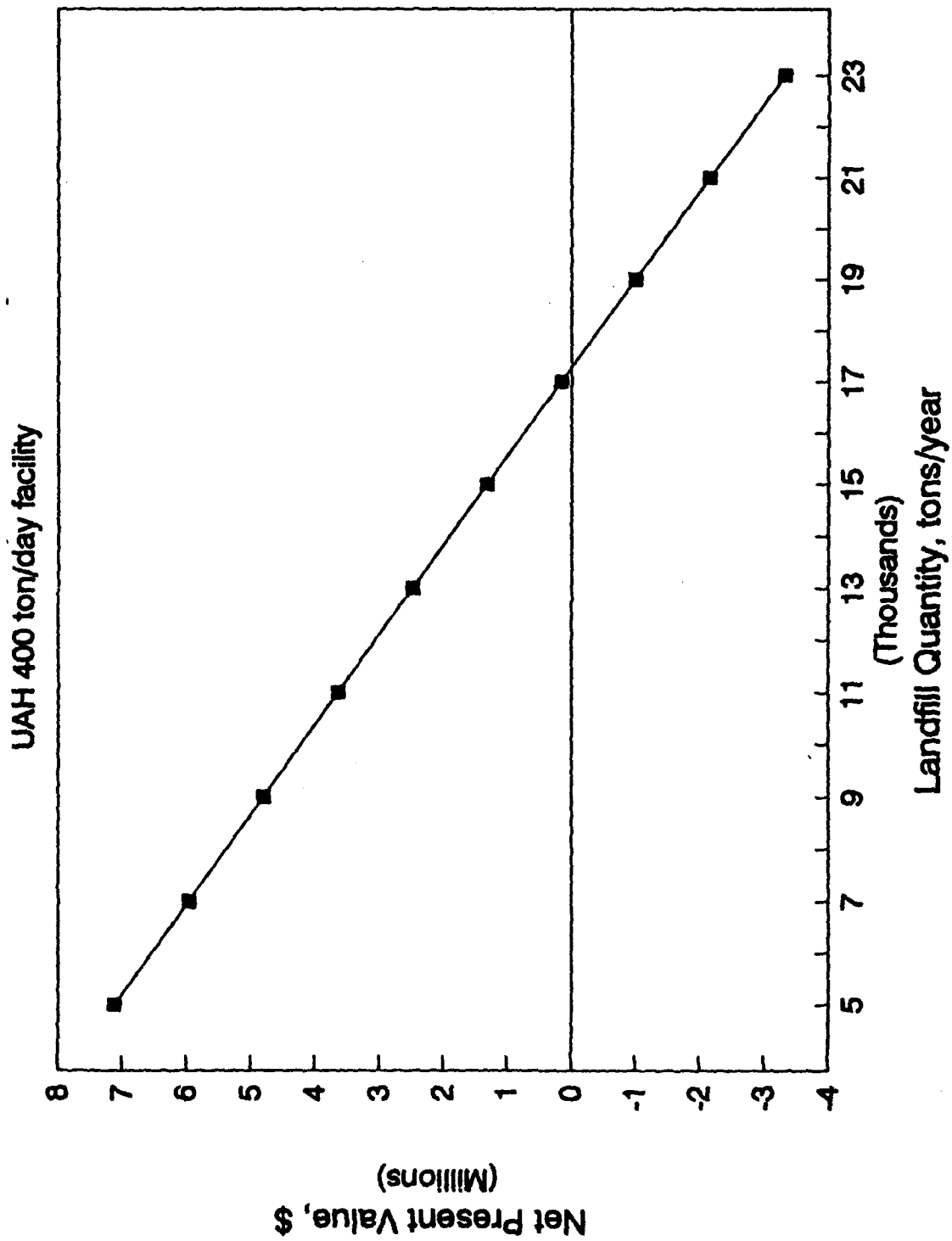


Figure 23. Sensitivity of Net Present Value to Landfill Quantity (UAH).

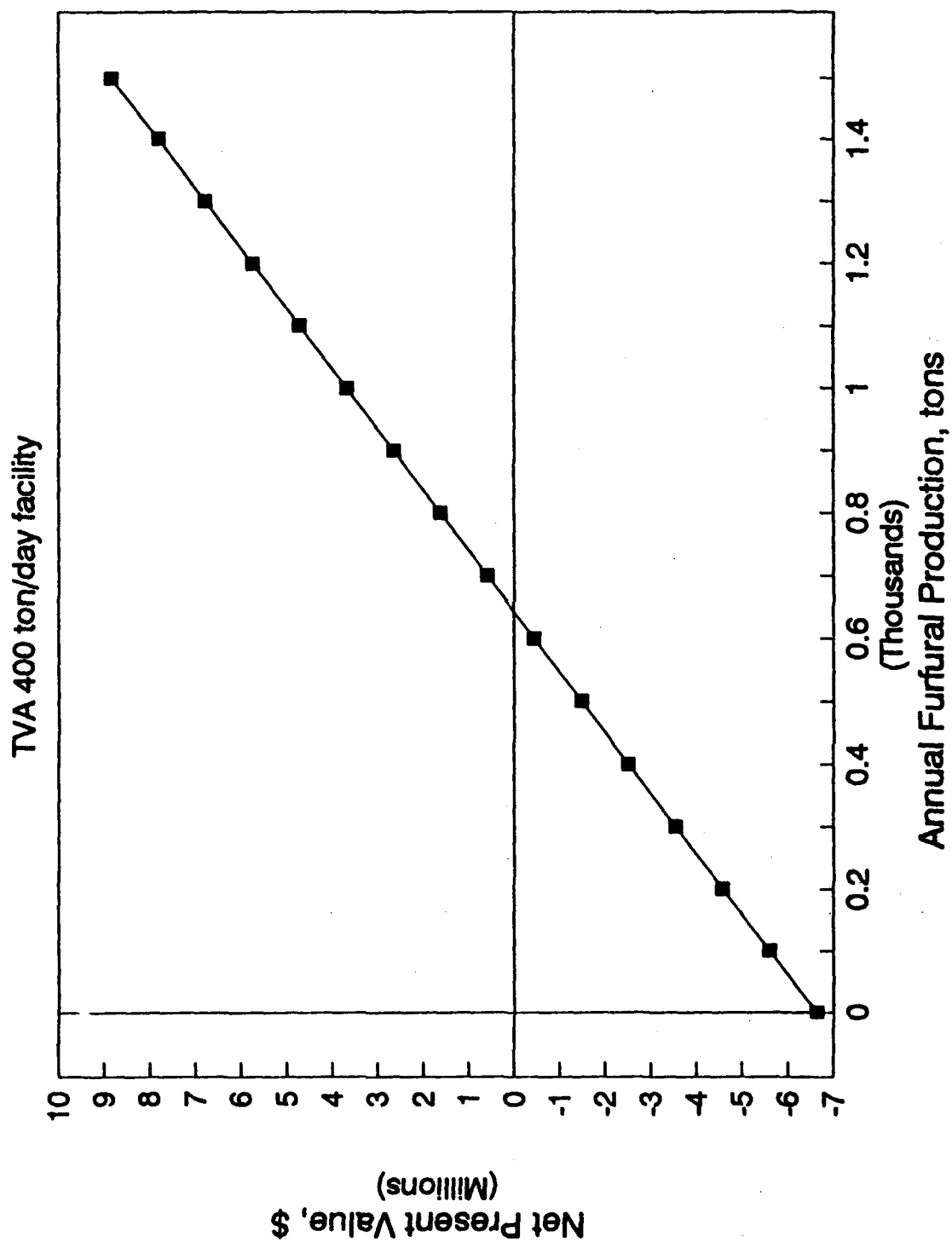


Figure 24. Sensitivity of Net Present Value to Furfural Production (TVA).



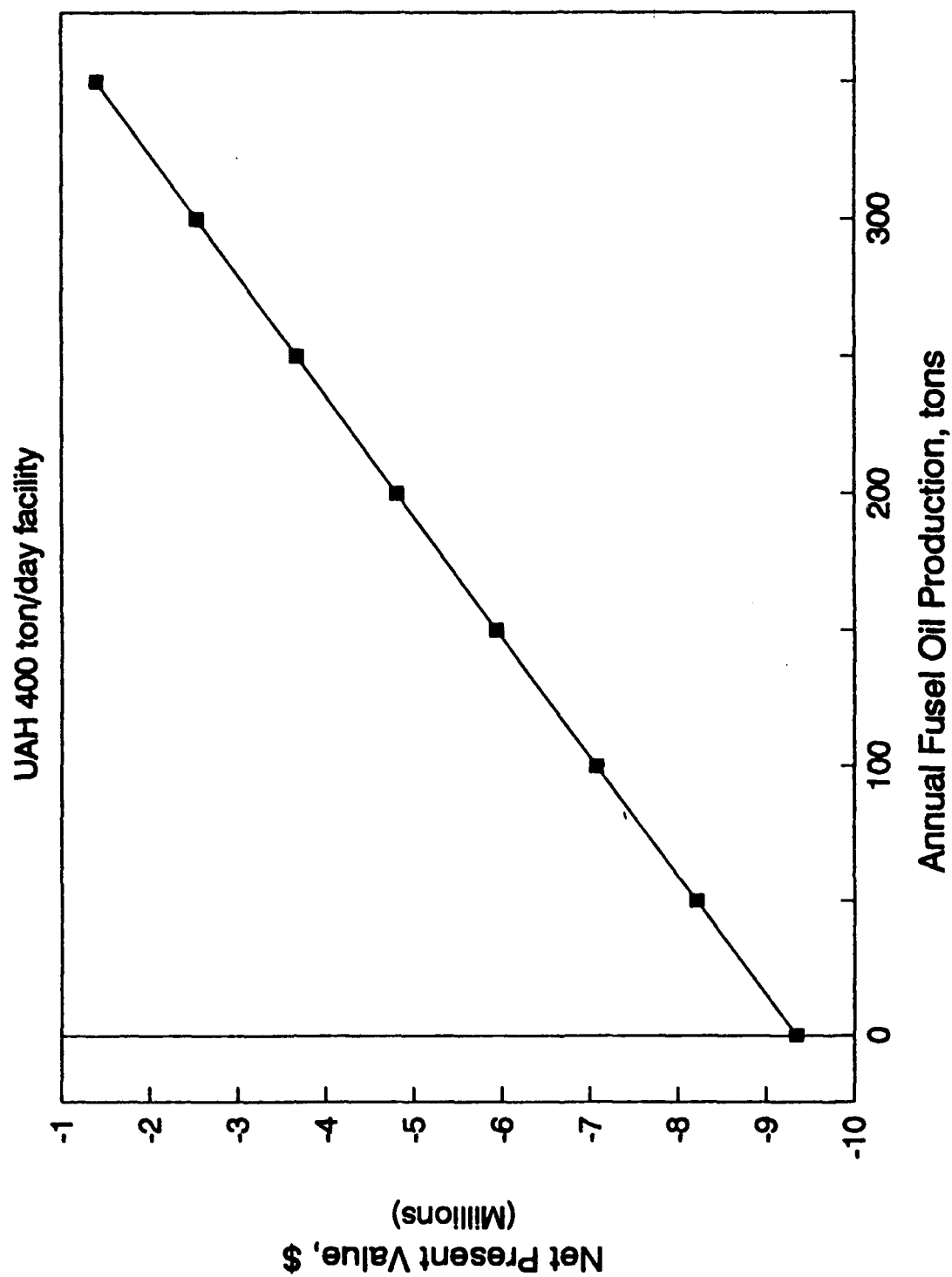


Figure 25. Sensitivity of Net Present Value to Fusel Oil Production (UAH).

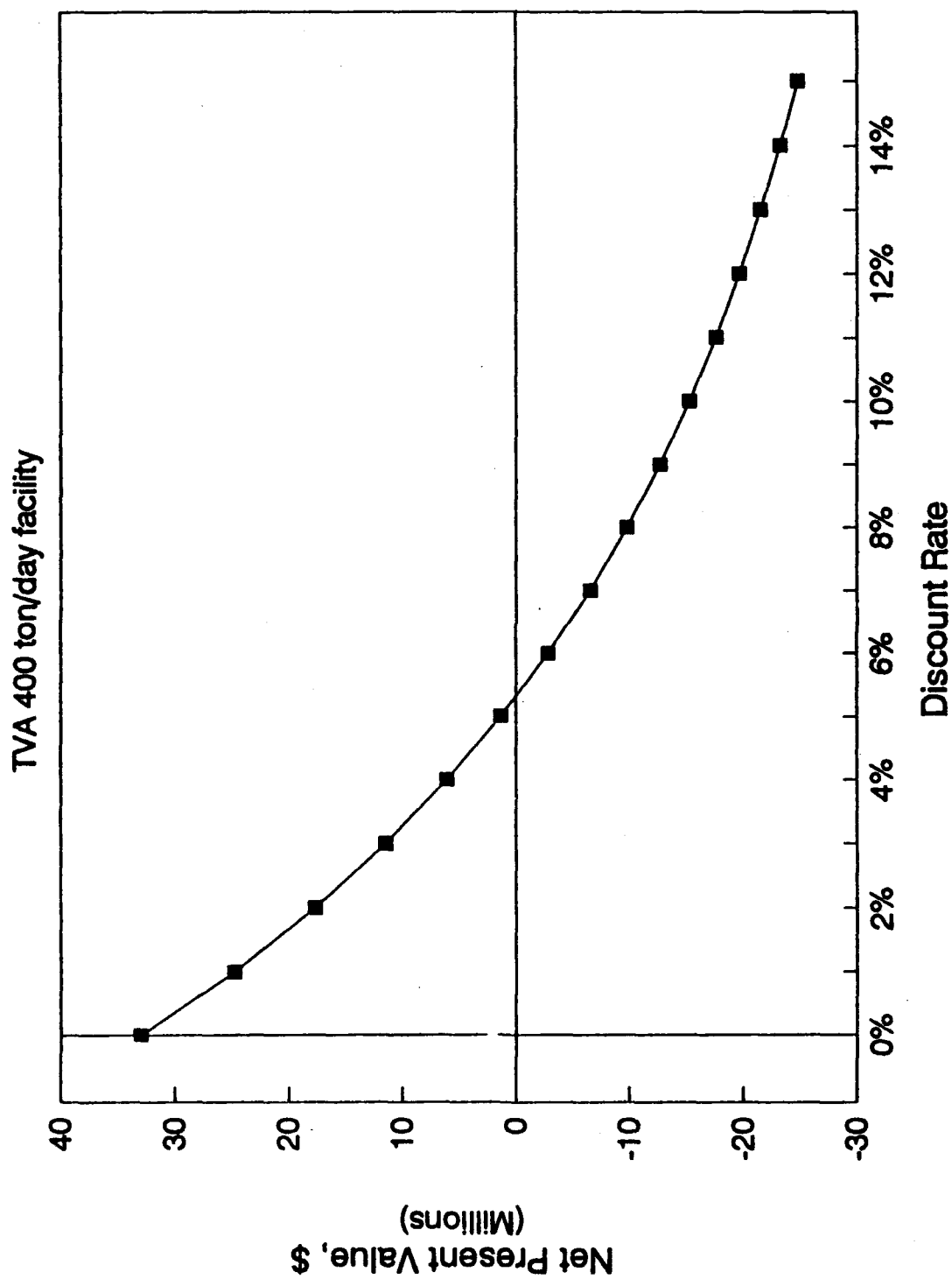


Figure 26. Sensitivity of Net Present Value to Discount Rate (TVA).

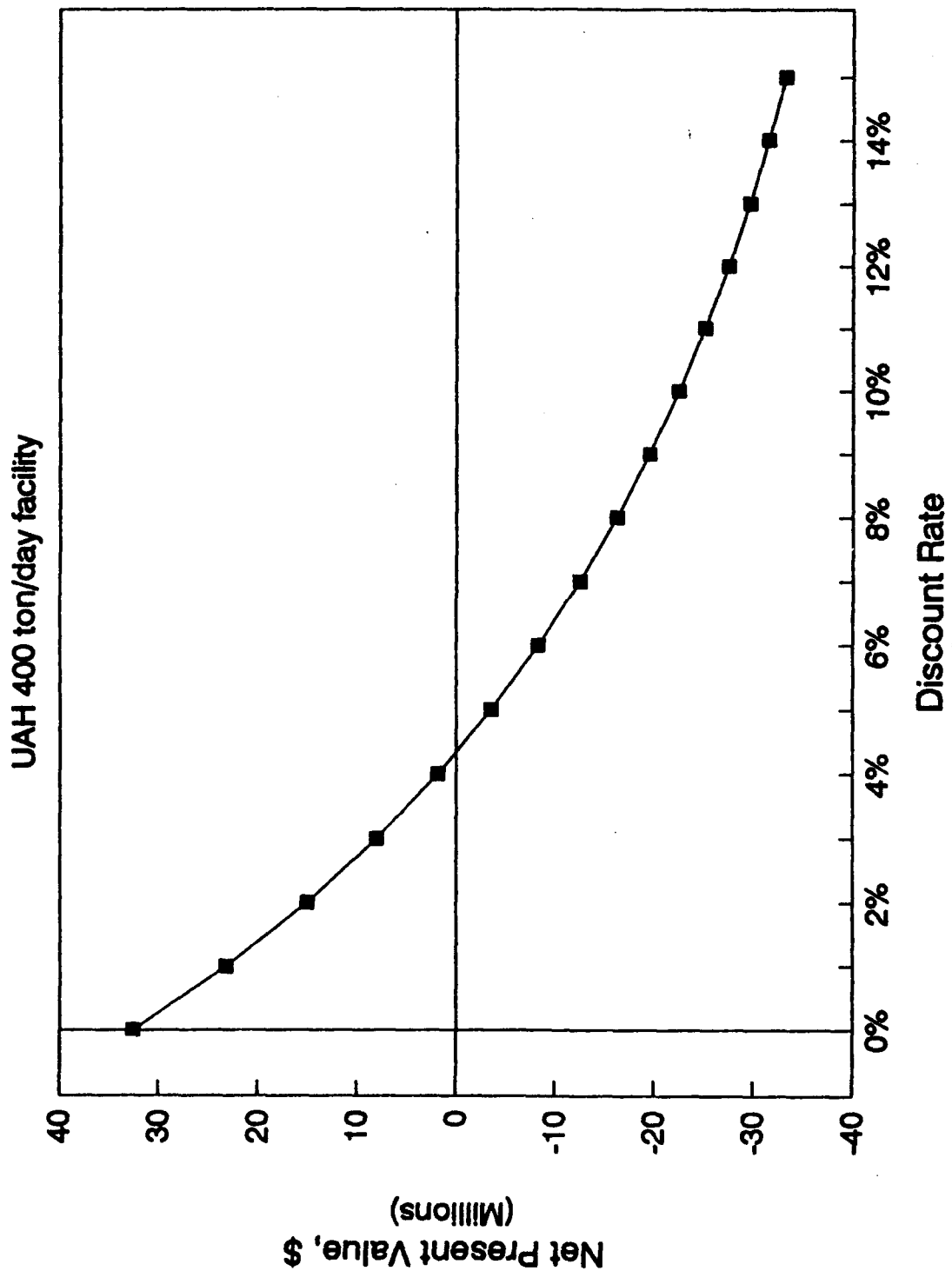


Figure 27. Sensitivity of Net Present Value to Discount Rate (UAH).

## 6 LITERATURE REVIEW

### Summary

Researchers conducted a computer data base search of the literature to ascertain the level of maturity of competing processes for the production of ethanol from paper, paper waste and other cellulosic materials (i.e., wood) via acid and enzymatic hydrolysis. Information on the yields of ethanol and byproducts was also sought to compare with those of the developers. Product and byproduct market information and current economic data were also obtained. A separate search yielded references to the criteria used for selecting plant sizes for MSW processing. The data bases searched included Paperchem, Chem Abstracts, Biotechnology Abstracts, and NTIS.

The search was largely limited to the last 10 years (1981 through 1991), although some earlier references were considered for background. A total of 178 references were cited. Relevant articles and papers were also obtained from private collections held by IGT staff, among other sources. Several additional references were obtained during the tours at the TVA (Muscle Shoals) and UAH (Huntsville) research facilities conducted on 23 and 24 April 1991. The list of references that resulted from the search is presented in the reference section of this report. Some articles were not directly relevant to the project and were, therefore, not reviewed. Also, more recently published articles that included economic evaluations were reviewed, while somewhat older references were not. Short synopses of the literature (augmented abstracts) that have been reviewed are included below. Copies of the articles and reports collected during the project have been appended. Many of these references are of interest to the ethanol evaluation program. A total of 76 references are in hand and have been read, reviewed, or otherwise considered for review. Of these references, many were prepared (or cowritten) by TVA, a few by UAH, and some by the University of Arkansas, Oak Ridge National Laboratory, Solar Energy Research Institute (SERI), and the University of Toronto.

The conversion of biomass and other cellulosic wastes to ethanol has been and continues to be the subject of many research efforts. Most of the processes studied can be divided into two general categories: (1) acid hydrolysis, and (2) enzymatic hydrolysis. Research has also been conducted to determine which modes of pretreatment can render the cellulosic wastes more easily and readily hydrolyzable. Similarly, researchers have also attempted to genetically engineer yeast (and other microorganisms) to readily convert the 5- and 6-carbon sugars produced by hydrolysis to ethanol without suffering the reduction in ethanol production that typically results from byproduct inhibition.

### Article Synopses

The references describe laboratory and somewhat larger-scale batch experimental testing of the processes being developed. Some are oriented toward improving the fermentation step by genetic engineering. Others have studied the hydrolysis of cellulose by different types of acids. None of the references report results on a scale comparable to that of TVA's 2 ton/day batch apparatus.

The synopses are arranged according to the organization that submitted or wrote the article—Tennessee Valley Authority, University of Alabama in Huntsville, University of Arkansas, Oak Ridge National Laboratory, Solar Energy Research Institute, University of Toronto, and then others alphabetically. The articles under each heading are arranged in alphabetical order according to the primary author's name.

## Tennessee Valley Authority

Barrier et al.<sup>25</sup> at the Tennessee Valley Authority, reported (1991) on pilot scale batch hydrolysis tests conducted with RDF, newsprint, and cardboard pellets. The tests were conducted in two different reactors:

1. The primary hydrolysis reactor (Sunds Defibrator, Inc.) is a cylindrical vertical reactor capable of operating at up to 400 psig and 215 °C (419 °F). The feed material is fed by compression screw into the reactor. The nominal feed rate is 2 ton/day (168 lb/h); however, equipment limitations resulted in RDF feed rates of about 100 lb/h. Twin parallel screws convey the feed upward and dilute H<sub>2</sub>SO<sub>4</sub> (about 2 wt percent) is injected into the reactor where the feed enters the twin screws. The material flows by gravity over a baffle into the main reactor.

2. The second hydrolysis reactor is a horizontal 21-in. diameter screw conveyor (Kamyr, Inc.) capable of operating at 235 psig and 215 °C with H<sub>2</sub>SO<sub>4</sub> concentrations up to 5 wt percent. Material is first presteamed then fed to the reactor concurrently with acid. Both reactors are constructed of zirconium to withstand acid attack.

Estimates of the alcohol yields from the RDF tests were not determined, because these were materials handling tests. The tests with cardboard pellets were conducted at 160 to 175 °C (320 to 347 °F), reaction times of 15 to 25 minutes, and H<sub>2</sub>SO<sub>4</sub> concentrations in the liquid of 1.8 to 2.0 wt percent. The ethanol yields (based on sugar content) ranged from 16 to 29 gal/ton. The tests with newsprint pellets were conducted at 160 °C, reaction times of 10 to 20 minutes, and H<sub>2</sub>SO<sub>4</sub> concentrations in the liquid of 2.0 to 2.5 wt percent. The ethanol yields ranged from 16 to 23 gal/ton. The authors indicate that improved sugar yields may be possible at higher temperatures (180 to 200 °C [347 to 392 °F]).

In another article by TVA, Barrier and Bulls<sup>26</sup> (1991) quantify the availability of biomass and wastes for feedstock. Biomass, in the form of wood and wood wastes, crop residues, and municipal solid wastes with average energy contents of about 7000 Btu/lb (dry basis) could make a significant impact on the U.S. energy picture. Two TVA processes are: (1) dilute two stage acid (H<sub>2</sub>SO<sub>4</sub>) hydrolysis conducted at 160 °C and 10 atmospheres and tested in a 2 ton/day facility, and (2) concentrated two-stage acid hydrolysis conducted at 100 °C (212 °F) and 1 atmosphere and tested in a 4 ton/day facility. Preliminary laboratory evaluations of waste-derived feedstock and newsprint have resulted in sugar yields equivalent to 25 to 40 gal (of ethanol)/ton of feedstock processed. Pilot scale evaluations of dilute acid hydrolysis of MSW have begun with results comparable to those achieved in the laboratory.

Barrier et al.<sup>27</sup> (1990) also describe the preliminary economics of producing alcohol and other coproducts (furfural, acetic acid, electricity) by dilute acid hydrolysis of waste derived feedstock (WDF). The total capital investment for the base case commercial facility-sized to process 500 ton/day of WDF, was estimated to be \$45 million. The return on investment (ROI) for the base case (\$15 tipping fee, 30 gal/ton ethanol, \$1.25/gal ethanol selling price) was 19.35 percent. The sensitivity of the ROI to tipping fee, price of electricity, and plant size were also estimated.

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<sup>25</sup> J.W. Barrier, M.M. Bulls, and G.E. Farina, "Pilot-Plant Evaluations of Dilute Acid Hydrolysis of Municipal Solid Waste," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

<sup>26</sup> J.W. Barrier, and M.M. Bulls, "Feedstock Availability of Biomass and Wastes," ACS Symposium Book Series *Emerging Materials and Chemicals From Biomass* (March 1991).

<sup>27</sup> J.W. Barrier, M.M. Bulls, J.D. Broder, and R.O. Lambert, "Production of Ethanol and Co-Products From MSW-Derived Cellulosics Using Dilute Sulfuric Acid Hydrolysis," *12th Symposium on Biotechnology for Fuels and Chemicals* (Gatlinburg, TN, 7-11 May 1990).

In another study by TVA, Barrier et al.<sup>28</sup> (1989) considered the potential contribution of a higher value byproduct, furfural, to the overall economics of producing ethanol from hardwood. Two cases were evaluated: concentrated acid hydrolysis and dilute acid hydrolysis. For the base case plant size of 500 tons of hardwood processed/day, the total plant investment (TPI) and ethanol production costs were estimated to be \$56.5 million and \$1.69/gal for the concentrated acid hydrolysis process and \$80.8 million and \$1.81/gal for the dilute acid hydrolysis process. When enhanced furfural production was included in the economics, the TPI and ethanol production costs were \$65.0 million and \$0.96/gal for the concentrated acid hydrolysis process, and \$67.7 million and \$0.64/gal for the dilute acid hydrolysis process. According to the authors, the domestic demand for furfural is declining due to lower activities in certain industries. However, reduced costs of producing furfural via acid hydrolysis may spur market growth. If that occurs, selling furfural as a byproduct could make ethanol production from TVA's acid hydrolysis processes cost competitive with other sources.

Broder and Barrier<sup>29</sup> (1988) described TVA's integrated biomass refining system, which can include the following byproducts: protein, lignin products, chemicals, ethanol, methane, single cell protein, aquaculture, and distiller's solids. They also provide a list of feedstocks tested by TVA in laboratory scale experiments, including conversions of hemicellulose and cellulose to xylose and glucose, respectively. TVA has studied the fermentation of xylose to ethanol using *Pachysolen tannophilus* as a standard. The conversion of xylose to ethanol is reportedly at only about 50 percent. The authors state that "... because commercial equipment is used in the plant (the 2 ton/day pilot plant), process guarantees can be obtained from the equipment vendors and the equipment can be scaled up to commercial size from pilot experiments." The TVA biomass processing system can be flexibly assembled, depending upon economics.

Bulls et al.<sup>30</sup> (1991) compare the economics of different MSW use technologies—ethanol production, RDF combustion, and mass burning—for a 1000 ton/day MSW plant. Ethanol production is based on TVA's dilute H<sub>2</sub>SO<sub>4</sub> process, which is described in brief (pilot scale results are listed as 22 to 29 gal of ethanol/ton of RDF). A preliminary economic evaluation by TVA determined the total plant investment for ethanol production, RDF combustion, and mass burning to be \$104 million, \$131 million, and \$105 million, respectively. Operating costs were estimated to be \$18.1 million, \$10.2 million, and \$7.9 million, respectively. Tipping fees were calculated to be \$38.37/ton for ethanol production, \$61.10/ton for RDF combustion, and \$56.67/ton for mass burning.

Fein et al.<sup>31</sup> (1991) describe the combined efforts of Bio-Hol (Canada), the Swedish Ethanol Development Foundation, and TVA to develop a wood-to-fuel ethanol process. The process is comprised of prehydrolysis, hydrolysis, and fermentation. In the prehydrolysis stage, the hemicellulose is converted to xylose with sulfur dioxide (2.5 wt percent) at 160 °C and a residence time of 30 minutes. Yields of 90 percent of theoretical were consistently achieved. Prehydrolysis was conducted in the Sunds Defibrator at the TVA Muscle Shoals facility, which was operated continuously for at least 100 hours for each production run.

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<sup>28</sup> J.W. Barrier, R.O. Lambert, and M.M. Bulls, "Potential Contribution of Furfural to the Economics of Ethanol Production From Hardwoods," *Southern Biomass Conference* (Blacksburg, VA, 12-14 September 1989).

<sup>29</sup> J.D. Broder and J.W. Barrier, "Producing Ethanol and Co-Products From Multiple Feedstocks," *Am. Soc. of Agricultural Engineers Summer Meeting* (Rapid City, SD, 26-29 June 1988).

<sup>30</sup> M.M. Bulls, et al., "Comparison of MSW Utilization Technologies—Ethanol Production, RDF Combustion, and Mass Burning."

<sup>31</sup> J.E. Fein, D. Potts, D. Good, M. Beaven, A. O'Boyle, D. Dahlgren, J.M. Beck, and R.L. Griffith, "Development of an Optimal Wood-to-Fuel Ethanol Process Using Best Available Technology," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

Hydrolysis of the pretreated wood was conducted in four triple pass heat exchangers in series (1-in. diameter, plug flow). The first three heat exchangers were used to preheat the slurry (15 to 20 percent solids). Hydrochloric acid was injected upstream of the fourth heat exchanger at 1 to 3 percent (dry solids basis). Maximum glucose yields were obtained at temperatures of 225 to 240 °C (437 to 464 °F) at 10 to 30 seconds with HCl. Glucose yields for this stage exceeded 50 percent. The authors indicate that both the prehydrolysis and hydrolysis reactors are commercially available.

Ten different strains of *Saccharomyces cerevisiae* were tested for suitability to ferment the hydrolyzate. Three methods were tested to remove fermentation inhibitors from the hydrolyzate before fermentation, including CaO/sulfite, activated carbon, and ion exchange. CaO/sulfite improved the fermentability of the hydrolyzate appreciably. An economic evaluation of the process showed ethanol production costs ranged from \$4.50 to \$1.36/gal.

Strickland et al.<sup>32</sup> (1987) described the process being developed by TVA to produce both pentose and hexose sugars derived from wood and convert them to alcohol. Sugar production occurs in a two-stage dilute sulfuric acid process to sequentially hydrolyze hemicellulose (enriched in xylose) and cellulose (enriched in glucose) with intermediate washing and re-acidification steps and differing hydrolysis conditions. Two interrelated hydrolysis approaches are being studied: high-temperature/short-residence time, and moderate-temperature/moderate-residence time. About 95 percent of the hemicellulose fraction of the wood is dissolved to produce sugars and other organics. Significant progress has been made in the laboratory on the conversion of hemicellulosic hydrolyzates to ethanol through pretreatment, nutrient additions, and control of fermentation conditions using the yeast *Pachysolen tannophilus*. Conversion of these hydrolyzates would significantly enhance the yield and economics of ethanol production from wood. Experience indicates moderate difficulty in the fermentation of cellulosic hydrolyzates. TVA estimated the cost to produce ethanol to be \$1.75/gal for a 1000 ton/day wood processing plant.

#### *University of Alabama (Huntsville)*

Coleman et al.<sup>33</sup> (1987) describe the patented Holloway Process as it relates to the pretreatment of MSW and sewage sludge for ethanol production. (This process was specified by UAH for pretreating the waste paper in their proposal.) The Holloway process equipment is described in some detail, including a schematic diagram. The unit is 15 ft long and 51 in. in diameter. The MSW basket rotates at 10 rpm. The overall ethanol yield was about 40 gal/ton of MSW (as received). This study considered both acid and enzymatic hydrolysis processes, but selected concentrated acid (HCl) for the economic evaluation. The economic evaluation of the process to convert MSW and sewage sludge to ethanol was based on scaling up a 500 ton/day plant to 1500 ton/day for a nonspecific site in Birmingham, AL. The total plant investment was estimated to be \$90 million and the break-even ethanol product cost was \$1.50/gal. Byproduct credits for tipping fees and recyclables would reduce the cost further.

Eley and Holloway<sup>34</sup> (1987) describe the Holloway process developed at UAH for sterilizing MSW feedstock prior to processing. The prototype is a cylindrical vessel about 5 ft in diameter and 12.5 ft long inclined about 15 degrees downward toward the exit door. It has means for agitating the contents. A typical batch charge consists of about 600 lb of MSW mixed with about 300 lb of water. Once the unit

<sup>32</sup> R.C. Strickland, R.L. Griffith, M.J. Beck, and J.R. Watson, "Conversion of Hardwoods to Ethanol—The Tennessee Valley Authority Approach," *Energy From Biomass and Wastes 11* (16-20 March 1987), Proceedings pp 981-996.

<sup>33</sup> D.R. Coleman, T.J. Laughlin, M.V. Kilgore, Jr., C.L. Lishawa, W.E. Meyers, and M.H. Eley, "The Bioconversion of Municipal Solid Waste and Sewage Sludge to Ethanol," *Biotechnol. Advan. Process. Munic. Wastes Fuels Chem.* (1987), pp 407-17.

<sup>34</sup> M.H. Eley and C. C. Holloway, "Treatment of Municipal Solid Wastes by Steam Classification for Recycling and Biomass Utilization," *Applied Biochem. and Biotech.* (1988), Vol 18, pp 125-135.

is loaded and sealed, sufficient steam (60 psig) is injected to heat the MSW to about 300 °F. The material is cooked for another hour after attaining the nominal temperature. The entire batch process requires about 3 hours to complete. The authors found that increasing the cooking time from 15 to 60 minutes increased the fraction of total product in the fines category (-1.3 cm) from 58 to 72 percent. At the same time, the mid fraction (-1.3 + 5.1 cm) decreased from 35 to 21 percent. An energy analysis was also included in the article.

*University of Arkansas (Fayetteville)*

Ackerson et al.<sup>35</sup> at the University of Arkansas (1991), studied the hydrolysis of the lignocellulosic fraction of MSW with concentrated HCl and the subsequent fermentation of the sugars to alcohol. The best conditions were found to be a high acid concentration (80 percent H<sub>2</sub>SO<sub>4</sub> or 41 percent HCl) and relatively mild temperature (≈40 °C [104 °F]). Near complete conversions of hemicellulose and cellulose were achieved. In batch fermentation tests with *Saccharomyces cerevisiae*, almost total (97.5 percent) conversion of sugars was obtained in 16 hours. The study recommended continuous (not batch) alcohol fermentation and conducted laboratory tests with columns of immobilized *S. cerevisiae*. The productivities (99 percent conversion) were about an order of magnitude greater than a CSTR and 60 times more than the batch reactor. Further, alcohol inhibition was reduced. The economic projection indicated a \$35 million capital cost for a 20 million gal/year plant.

The University of Arkansas is developing<sup>36</sup> (1982) a process for converting municipal wastes, forestry residues, and other products to ethanol. The process includes: (1) feedstock preparation; (2) production of enzymes from a mutant strain of *Trichoderma reesei*; (3) production of a mutant yeast strain of *Candida brassicae*; (4) production of alcohol using simultaneous saccharification and fermentation (SSF); and (5) separation and concentration of products and byproducts. Tests were conducted in a 1 ton/day (feedstock) pilot plant in both batch and continuous modes. The pilot plant represented a 100 fold increase in size from the laboratory tests. The results of the pilot scale work compared favorably with, and in some instances surpassed, those from the laboratory. The use of unpasteurized feedstock resulted in lower yields in the SSF step due to feedstock competition by contaminating organisms.

The economic evaluation was conducted on a plant size of 2000 ton/day (oven dry), which was determined to be near optimum. The authors estimated that the minimum economical plant size would be near 1000 ton/day. For the 2000 ton/day plant, the TPI was estimated to be \$165.2 million. The selling price for alcohol required for a 15 percent after tax ROI (80 percent bond financing) was estimated to be \$1.59/gal. A private company, operating under a license from the University of Arkansas Foundation, sought to raise funds for a 50 ton/day MSW (75 percent) and pulp mill waste (25 percent) demonstration plant to resolve specific questions concerning feedstock pretreatment and animal feed byproduct testing. (The status of this effort is unknown.)

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<sup>35</sup> M.D. Ackerson, E.C. Clausen, and J.L. Gaddy, "Production of Ethanol From MSW via Concentrated Acid Hydrolysis of the Lignocellulosic Fraction," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

<sup>36</sup> K.J. Bevernitz, S.J. Gracheck, D.B. Rivers, D.K. Becker, K.F. Kaupisch, and G.H. Emert, "Development of Enzyme-Catalyzed Cellulose Hydrolysis Process for Ethanol Production," *Energy From Biomass and Wastes 6* (Lake Buena Vista, FL, 25-29 January 1982), Proceedings pp 897-918.



Clausen and Gaddy<sup>37</sup> (1988) describe a process under development at the University of Arkansas for ethanol production from corn stover comprised of:

1. Feedstock preparation by grinding the corn stover to -20 mesh to give the maximum possible slurry concentration
2. Two-stage acid hydrolysis involving hemicellulose removal by prehydrolysis with a low acid concentration (2 N H<sub>2</sub>SO<sub>4</sub>) under mild conditions (100 °C), followed by cellulose breakdown using 14 N acid at ambient temperature and avoiding toxic sugar/degradation product formation
3. Ethanol fermentation using *Saccharomyces cerevisiae* in batch culture with added yeast extract since the corn hydrolyzate served only as a carbon source
4. Acid recovery by filtration and countercurrent extraction. The total cost plus profit for alcohol production using this process was calculated at \$1.85/gal for a facility converting corn stover into 20 million gal of ethanol/year. One detrimental factor to the economics of the process was the very dilute solutions that resulted from acid hydrolysis. The process could be improved by solids concentration, acid recycling, and continuous fermentation.

#### *Oak Ridge National Laboratory*

Bienkowski et al.<sup>38</sup> (1987) described the various processing steps involved in the production and purification of antibiotics. The production of antibiotics is divided into fermentation and separation/purification. The separation and purification section of an antibiotic plant can be very large as a result of the number of processing steps required (up to 60) and the need to purify and recycle large quantities of organic solvents. Research is in progress on three relatively new separation techniques, specifically conventional and annular chromatography, supercritical extraction, and various membrane processes.

Byers<sup>39</sup> (1987) described an integrated process in which woody biomass was converted to fuel and simulated using the ASPEN simulator. The process was divided into four sections: (1) biomass refining, (2) hydrolysis, (3) bioreaction, and (4) fuel separation. Detailed attention was paid to the hydrolysis process and the bioreactor, while a general simulation of the biomass refining process and fuel separation step is included. A butanol fermentation using *Clostridium acetobutylicum* is shown as the fuel conversion process. A simulation of the biomass and active microbial system required the definition of non-conventional streams. A series of studies was performed in which the acid recycle ratios (the ratio of recycled acid to makeup acid in the reactor) varied from zero to an average of 0.75. As the recycle ratio increased, the operating cost of the overall process passed through a minimum. Suggestions for the refinement and extension of this approach are discussed. Its advantages in establishing the cost of proposed technologies, assessing areas where research and development are required and evaluating schemes for enhancing energy efficiency were evaluated.

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<sup>37</sup> E.C. Clausen and J.L. Gaddy, "Ethanol From Biomass by Concentrated Acid Hydrolysis and Fermentation," *Energy From Biomass and Wastes 12* (New Orleans, LA, 1988), Proceedings pp 1319-1342.

<sup>38</sup> P.R. Bienkowski, D.D. Lee, and C.H. Byers, "Evaluation of Separation and Purification Processes in the Antibiotic Industry," *Appl. Biochem. Biotechnol.* (1988), Vol 18, pp 261-273.

<sup>39</sup> C.H. Byers, "An Aspen Simulation of Fuel Production by Hydrolysis of Woody Biomass—Using *Clostridium acetobutylicum* Butanol Fermentation as Model," *Appl. Biochem. Biotechnol.* (1988), Vol 18, pp 143-57.

Bergeron et al.<sup>40</sup> (1988) present several dilute acid hydrolysis processes for conversion of lignocellulosic biomass into ethanol. Biomass composition and dilute acid hydrolysis kinetics are briefly reviewed to establish the limitations on dilute acid catalyzed breakdown of cellulose and hemicellulose. Descriptions are given for three dilute acid catalyzed processes using the plug flow reactor, the percolation reactor, and the progressing batch reactor, respectively. The economics of each option are summarized along with projections of possible cost improvements. The plug flow and progressing batch reactor processes may have commercial viability. High yields have been achieved with the plug flow reactor at wood concentrations below 12 percent, and experiments were initiated with a prehydrolysis reactor to allow processing of higher wood levels (18 to 20 percent). A tandem reactor configuration of the progressing batch reactor was in operation in February 1988 to evaluate the potential of achieving the high yields and sugar concentrations of countercurrent operation.

According to a paper by Bull<sup>41</sup> (1989), research on acid hydrolysis processes for converting cellulosic materials to ethanol is being phased out, while enzymatic processes appeared to be more promising. Most efforts have been focused on simultaneous saccharification and fermentation as a means to reduce endproduct inhibition of the enzymes. In the area of xylose fermentation, Bull reported that TVA researchers, using *P. tannophilus*, have achieved 2.0 to 2.5 percent ethanol concentrations with yields 70 percent of maximum theoretical capacity. Substantial progress has been made in all phases of research in the past few years. The predicted cost to produce ethanol from biomass in the late 1970s was \$3.60/gal. Current predictions show a cost of about \$1.35/gal with the potential to be reduced to less than \$0.80/gal.

Spindler et al.<sup>42</sup> (1991) conducted laboratory scale tests with different feedstocks, including corn cobs and stover, investigating simultaneous saccharification and fermentation with *Saccharomyces cerevisiae* and *Brettanomyces clausenii* alone and in mixed cultures. The results showed that the overall rates and yields were better in the mixed culture than *S. cerevisiae* alone, because of the additional  $\beta$ -glucosidase activity associated with *B. clausenii*. However, *S. cerevisiae* performed slightly better when substantial  $\beta$ -glucosidase was added, apparently because it has a higher ethanol tolerance. These results point out the necessity of providing high  $\beta$ -glucosidase levels to prevent the accumulation of the strong inhibitor, cellobiose (disaccharide of glucose) in the solution.

Wright<sup>43</sup> (1988) described current research into ethanol and liquid fuel production from lignocellulosic biomass, specifically cellulose hydrolysis, xylose fermentation and lignin conversion. Enzymatic hydrolysis was discussed with reference to: (1) pretreatment to make the biomass more easily digestible by enzymes and to separate a lignin fraction for conversion to liquid fuel; (2) enzyme production; (3) separate hydrolysis and fermentation; and (4) simultaneous saccharification and fermentation (SSF). Research into xylose fermentation includes possible integration of xylose fermentation with SSF processing. Potential xylose fermentation systems include *P. tannophilus*, *Candida shehatae* and *Pichia stipitis*. Research into lignin conversion has indicated the potential of chemical conversion of lignin to methyl aryl ethers (MAE), which have a high octane value. Environmentally related costs account for

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<sup>40</sup> P.W. Bergeron, J.D. Wright, and C.E. Wyman, "Dilute Acid Hydrolysis of Biomass for Ethanol Production," *Energy From Biomass and Wastes 12* (New Orleans, LA, 15-19 February 1988), Proceedings pp 1277-1296.

<sup>41</sup> S.R. Bull, "Advances in Processes for Fermentation Ethanol," *Energy From Biomass and Wastes 13* (New Orleans, LA, 13-17 February 1989).

<sup>42</sup> D. Spindler, C. Wyman, and K. Grohmann, "Ethanol Production by Simultaneous Saccharification and Fermentation (SSF) of Pretreated Woody Crops, Herbaceous Crops, Corn Cobs, and Corn Stover," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

<sup>43</sup> J.D. Wright, "Evaluation of Enzymatic Hydrolysis Processes," *Energy From Biomass and Wastes 12* (New Orleans, LA, 15-19 February 1988), pp 1247-1276.

about 10 percent of the total cost to produce ethanol from biomass. Ethanol and MAE could be produced at low cost using a process comprising SSF, xylose fermentation and lignin conversion.

Wright et al. (1987)<sup>44</sup> also described experiments with simultaneous saccharification and fermentation (SSF) using Genencor 150L cellulase and mixed yeast cultures of *Brettanomyces clausenii* and *S. cerevisiae*, which produced yields and concentrations of ethanol from cellulose of 80 and 4.5 percent, respectively. The system resulted in increased yields and product concentrations compared with the separate hydrolysis and fermentation process. The mixed culture was used because *B. clausenii* can ferment cellobiose (thus reducing endproduct inhibition), while *S. cerevisiae* ferments the monomeric sugars. Because yield was strongly dependent on reaction time, the analysis was carried out for a constant reaction time of 7 days. Initial cellulose concentration ranged from 7.5 to 15 percent, and enzyme loading varied from 7 to 13 IU/g cellulose. Yield was maximized at low concentrations and high enzyme loadings. Experimental results were combined with a process model to evaluate the economics of the process and to investigate the effect of alternative processes, conditions, and organisms. The cost to produce ethanol by the SSF process was estimated to be \$1.78/gal compared to \$2.66/gal for separate hydrolysis and fermentation process. The article discusses the processing steps in detail.

Another reference by Wright and Power<sup>45</sup> (1986) evaluated five major acid hydrolysis processes for the conversion of lignocellulosic feedstocks to ethanol. A complete conversion process was defined, and alternative hydrolysis processes integrated into this framework. The sensitivity of the process economics to technical improvements was also evaluated. In dilute acid hydrolysis processes, the acid hydrolyzes cellulose to glucose while simultaneously degrading the product sugars. The progressing batch reactor uses several percolation reactors in series to quickly remove the sugars from the reaction zone, minimizing sugar degradation while minimizing product dilution. The high temperature plug flow reactor has low yields, but converts the xylan fraction of the wood to furfural, a valuable chemical intermediate. The concentrated acid processes break down the crystalline structure of the cellulose, rendering it amorphous and readily hydrolyzable at conditions where sugar degradation is minimal and yields approach 100 percent. The process suffers from the expenses associated with handling large quantities of acid. The concentrated sulfuric acid processes consume large quantities of acid, and suffers from high acid replacement costs. The HCl processes recover the acid by distillation, which requires large amounts of energy and extremely expensive corrosion resistant equipment. The HF-based process has high acid consumption costs because HF is very expensive, and would suffer high capital and operating costs because of the extremely hazardous nature of the catalyst.

In general, dilute acid hydrolysis processes were preferred to concentrated acid processes. The increased ethanol yields afforded by concentrated acid processes were offset by increased costs incurred for using large quantities of acid among other factors.

#### *University of Toronto*

Lawford and Rousseau<sup>46</sup> (1991) at the University of Toronto conducted a laboratory investigation of the fermentation of xylose and glucose by a recombinant bacteria *Escherichia coli* carrying a plasmid with genes for pyruvate decarboxylase and alcohol dehydrogenase cloned from *Zymomonas*. The recombinant converted hemicellulose hydrolyzate (3.5 percent xylose) to ethanol at an efficiency of 94 percent.

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<sup>44</sup> J.D. Wright, C.E. Wyman, and K. Grohmann, "Simultaneous Saccharification and Fermentation of Lignocellulose: Process Evaluation," *Applied Biochem. and Biotech.* (1988), Vol 18, pp 75-90.

<sup>45</sup> J.D. Wright and A.J. Power, "Comparative Technical Evaluation of Acid Hydrolysis Processes for Conversion of Cellulose to Alcohol," *Energy From Biomass and Wastes 10* (Washington, DC, 7-11 April 1986), pp 949-971.

<sup>46</sup> H.G. Lawford and J.D. Rousseau, "Xylose to Ethanol: Enhanced Yield and Productivity Using Genetically Engineered *Escherichia coli*," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

A maximum volumetric ethanol productivity of 0.76 g/L-h was observed in  $\text{Ca}(\text{OH})_2$  treated prehydrolyzate prepared by the Bio-Hol Process. A comparison of these results with others demonstrates that this *E. coli* construct rivals the best of the current xylose fermenting organisms.

### Others

Aravamuthan et al.<sup>47</sup> at the University of Alabama (Tuscaloosa, 1988) conducted laboratory batch tests with hardwood mixtures (primarily red oak) to determine the effects of prehydrolysis treatment with sulfur dioxide followed by acid catalyzed ethanol/water delignification on subsequent cellulose hydrolysis. Typical acid hydrolysis causes condensation of the lignin polymer, which increases its molecular weight and may reduce the value of lignin based byproducts. Lignin condensation also reduces the susceptibility of cellulose glycosidic bonds to acid hydrolysis in the second stage (cellulose hydrolysis). Thus, removing lignin from the wood matrix without condensation could increase its byproduct value from that of a fuel ( $\approx \$0.04/\text{lb}$ ) to as much as  $\$0.30/\text{lb}$ . Also, the yield of glucose from the second stage of acid hydrolysis can be as much as 50 percent higher than if the lignin were not removed.

The results showed xylose recovery levels of up to 86 percent of the theoretical value. The optimum prehydrolysis conditions were  $\text{SO}_2$  concentrations of 0.25 to 0.50 percent, temperatures of 125 to 130 °C (257 to 266 °F), and 2 to 3 hours at temperature. The economic evaluation compared the base case two stage acid hydrolysis process (lignin as fuel) with  $\text{SO}_2$  prehydrolysis and acid catalyzed ethanol/waste delignification. The payout was sensitive to solvent recovery efficiency. With a solvent recovery of 99 percent, a  $\$0.30/\text{lb}$  credit for lignin, and increased ethanol yields, the payout period was reduced from the base case 2.6 years, to 1.4 years.

Berglund and Richardson<sup>48</sup> at Idaho National Engineering Laboratory (INEL) reported on the construction and operation of a small batch processing plant for converting corn (or other grain) to ethanol. The plant was sized to produce 100 L/h of 190 proof ethanol. The corn was ground to -80 mesh and then treated with  $\alpha$ -amylase for 1.5 hours in continuous plug flow reactors at about 200 °F.  $\alpha$ -Amylase converts the starch in the corn to dextrins. Next, the slurry was treated with gluco amylase, which converts the dextrins to glucose. Yeast was then added to the slurry to ferment the glucose to ethanol, a process designed to take 50 hours. The resulting beer mash contained 8 to 10 wt percent ethanol, which was distilled to 190 proof alcohol. The ethanol yield was 2.6 gal/bushel of corn. The equipment cost was  $\$509,359$  and installation was  $\$216,200$  (1982).

Borgwardt et al.<sup>49</sup> (1991) described the environmental benefits of the Hydrocarb process for converting biomass plus methane to methanol. Methanol would be used as a transportation fuel with carbon black as a byproduct. The preliminary economics of producing methanol from the Hydrocarb process are compared with those of other conventional processes for producing methanol or ethanol.

Bungay<sup>50</sup> (1982) described some of the processes that were being developed to convert biomass to fuel ethanol. He concluded that higher value products, such as lignin, molasses, and paper pulp, would be the focus of the first commercial factories rather than ethanol.

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<sup>47</sup> R. Aravamuthan, W.-Y. Chen, K. Zargarian, and G.C. April, "Ethanol From Southern Hardwoods: Role of Presulfonation in the Acid Hydrolysis Process," *Chemical Engineering Community* (1988), Vol 69, pp 115-127.

<sup>48</sup> G.R. Berglund and J.G. Richardson, "Design for a Small-Scale Fuel Alcohol Plant," *Chem. Eng. Prog.* (1982), Vol 78, No. 8, pp 60-67.

<sup>49</sup> R.H. Borgwardt, M. Steinberg, E.W. Grohse, and Y. Tung, "Biomass and Fossil Fuel to Methanol and Carbon Via the Hydrocarb Process," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

<sup>50</sup> H.R. Bungay, "Biomass Refining—Conversion of Lignocellulosic Materials to Fermentable Sugars: Importance of Lignin to the Profitability of the Process," *Science* (1982), Vol 218, No. 4573, pp 643-46.

In an article by Gaines and Karpuk<sup>51</sup> (1986), the markets for ethanol and co- or byproducts are discussed in detail. Ethanol is the primary product of biomass hydrolysis and fermentation processes. The quantity and quality of coproducts depend on the feedstock and on the process steps employed. When the feedstock is lignocellulosic—wood or herbaceous energy crops, such as napier grass or kenaf—substantial quantities of lignin and five carbon sugars and derivatives are produced, and the sale of these and other coproducts has a potentially large impact on the economics of fuel alcohol production. For each product, potential markets and materials competing for these markets (or competing routes to the same product) were identified. The value of ethanol as a chemical, octane enhancer, and fuel was estimated. Coproduct credits were calculated, and the potential benefits resulting from coproduct recovery were estimated. The quantity of each material that could be produced by fermentation was compared with the size of potential markets. Where coproducts of substantial fuel ethanol manufacture could saturate their markets, the question of whether early sales could serve as a wedge to ethanol market penetration was addressed. The chemical market for ethanol in 1984 was  $170 \times 10^6$  gal/year at a price of \$1.80/gal.

An article by Hocker<sup>52</sup> (1991) described the status of WTE projects in the United States and listed companies that are heavily involved in WTE projects. The average waste handling capacity of all WTE plants brought on line prior to 1988 was 574 ton/day. Recently, however, the average plant size has been 1056 ton/day. Some observers feel that economies of scale and return on investment warrant plant sizes in the 1000 to 1500 ton/day range. Others feel that future WTE plants will be in the 300 to 800 ton/day range, because that size is appropriate for many local communities. Also, because of community recycling efforts, the feedstock for the WTE plants will become more like that produced from plants that convert MSW to RDF. It should be emphasized that these are waste-processing plants, not energy plants.

Koukios and Sidiras<sup>53</sup> (1991) present the results of modeling studies on prehydrolysis of biomass and subsequent hydrolysis of the cellulosic fraction. Each step was assumed to follow pseudo-first-order kinetics. Prehydrolysis can be an effective and potentially efficient fractionation stage for hemicellulose. Further, hemicellulose fractionation is accompanied by partial depolymerization of cellulose and extensive depolymerization of lignin.

Lynd et al.<sup>54</sup> (1991) consider the prospects for fuel ethanol production from cellulosic biomass (including MSW) to be favorable. However, conversion economic factors must be overcome. About 1.25 gal of ethanol (neat) are required to travel the same distance as that obtained from 1 gal of gasoline in optimized engines. Thus, at the 1989 average wholesale gasoline price of \$0.655/gal, to be competitive, the selling price of unsubsidized ethanol must be 0.52/gal. By the year 2000, the wholesale price for ethanol must be \$0.70/gal to be competitive with \$0.88/gal gasoline. Environmental benefits include a low net emission of CO<sub>2</sub> to the atmosphere, among others. The cost of enzyme constrains the hydrolysis reaction time to values far above the limit imposed by substrate reactivity. Improvements that will reduce the cost to produce ethanol include increasing pretreated substrate reactivity, improving enzyme production systems, improving enzyme reactivity, and enzyme recycling. Research is likely to result in enzyme-based processes that are significantly cheaper than acid-based processes. The technology could be developed enough to produce ethanol on a cost competitive basis with gasoline by the year 2000.

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<sup>51</sup> L.L. Gaines and M. Karpuk, "Fermentation of Lignocellulosic Feedstocks: Product Markets and Values," *Energy From Biomass and Wastes 10* (Washington, DC, 7-10 April 1986), Proceedings pp 1395-1416.

<sup>52</sup> Hocker, p 37.

<sup>53</sup> E.G. Koukios and D.K. Sidiras, "Designing Biomass Refineries To Produce Fuels, Paper Pulp and Chemicals—The Key Role of Prehydrolysis," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

<sup>54</sup> Lynd et al, p 72.

Maiorella et al.<sup>55</sup> (1984) compared 11 alternative fermentation schemes for ethanol production including batch fermentation, simple continuous fermentation (continuous stirred tank reactor (CSTR)), CSTR with recycle, CSTRs in series (perforated plate tower fermenter), plug flow fermenters and immobilized cell tower fermenters. The article describes in considerable detail a fermentation kinetics model with a fermentation plant design model. The inhibitory effect of ethanol on the specific ethanol productivity of *Saccharomyces cerevisiae* is graphically presented. The fermentation processes were subjected to economic comparison and selective ethanol removal processes, e.g., membrane, extraction and vacuum processes, were discussed. The economics of ethanol production using cellulose derived sugars were evaluated. The cost of glucose obtained from corn stover hydrolysis was compared with that from molasses. The final ethanol product cost ranged from 43.94 to 53.05 ¢/l (\$1.66 to \$2.01/gal) depending upon the processing steps. A selective ethanol removal step during fermentation was also considered to reduce inhibition of the yeast.

Nystrom et al.<sup>56</sup> (1984) discuss the various advantages and disadvantages of acid and enzyme hydrolysis processes, and direct microbial conversion of cellulosic materials to ethanol. In the area of alcohol purification, molecular sieves, cellulose and starch adsorbents, and extractive distillation are listed. The authors also provide an economic comparison of hardwood conversion via these processes. The results show that the required selling price for ethanol ranged from \$0.17/liter (\$0.64/gal) with enzymatic hydrolysis when the lignin was sold as a phenolic extender, to \$0.86/l (\$3.26/gal) when the lignin was used as fuel. The selling price of alcohol from the dilute acid hydrolysis process was \$0.68/l (\$2.57/gal). None of the cellulose conversion processes can compete with conventional grain to ethanol technology.

Parsons et al.<sup>57</sup> (1991) conducted a systematic study to determine the most appropriate solvent for recovering ethanol from dilute aqueous solutions. After screening 1361 solvents using a computer-based ranking system, oleyl alcohol (*cis*-9-octadecen-1-ol) was selected. Continuous tests conducted with the alcohol showed it to be effective for ethanol extraction from dilute solution. An economic evaluation compared the cost of producing 95 percent ethanol from dilute aqueous solutions by the solvent extraction process or conventional distillation. For an ethanol concentration of 1 percent, the capital and energy costs were 5 percent and 71 percent lower, respectively, for the solvent extraction process than those for conventional distillation. For an ethanol concentration of 5 percent, the capital and energy costs were 24 percent and 15 percent lower for the solvent extraction process than those for conventional distillation.

Zerbe and Baker<sup>58</sup> (1986) discuss the fundamental analysis of ethanol production from southern red oak and make recommendations for a process that is technically ready for commercial exploitation, but requires some equipment testing and evaluation. Data for hemicellulose and cellulose hydrolysis were correlated by using models for developing and evaluating the process design. Coproducts originating from hemicellulose were assumed to be animal feed molasses, furfural, and acetic acid. The potential for the process could be significantly improved through utilization of the pentose sugars, which might be realized by fermentation to ethanol. The best existing design for using the hydrolyzate of the second stage is one in which all products other than glucose are concentrated and burned. The lignin residue could also be burned as fuel, but the process heat requirements are about 50 percent greater than could be supplied by the lignin. Advantages of the two stage process include high concentration of product solutions and

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<sup>55</sup> B.L. Maiorella, H.W. Blanch, and C.R. Wilke, "Economic Evaluation of Alternative Ethanol Fermentation Processes—Including Selective Ethanol Removal Systems," *Biotechnol. Bioeng.* (1984), Vol 26, No. 9, pp 1003-25.

<sup>56</sup> J.M. Nystrom, C.G. Greenwald, F.G. Harrison, and E.D. Gibson, "Making Ethanol From Cellulosics," *Chem. Eng. Progr.* (May 1984), Vol 80, No. 5, pp 68-74.

<sup>57</sup> R.V. Parsons, H. Husain, A.J. Daugulis, and S. Vihayan, "Extraction of Ethanol From Dilute Solutions," *Energy From Biomass and Wastes 15* (Washington, DC, 25-29 March 1991).

<sup>58</sup> J.I. Zerbe and A.J. Baker, "Investigation of Fundamentals of Two-Stage Dilute Sulfuric Acid Hydrolysis of Wood," *Energy From Biomass and Wastes 10* (Washington, DC, 7-10 April 1986), Proceedings pp 927-947.

somewhat reduced capital requirements compared to the percolation process, but these advantages are offset to some extent by higher yields from the percolation process.

References collected, but not specifically included in the literature review above are listed in the references section. (p 109)

## 7 CONCLUSIONS AND RECOMMENDATIONS

### Conclusions

Based on a review of feasibility studies prepared by TVA and UAH, the published literature, information gathered by visits to TVA and UAH facilities, and researchers' independent process scale up and economic evaluation, the following observations were made.

- Use of MSW through recovery of recyclable materials and conversion of the cellulosic fraction to both useful and valuable byproducts can significantly reduce the burden on existing landfills for waste disposal.
- The production of fuel ethanol from MSW will help reduce U.S. dependence on imported liquid fuels.
- The scale-up of these processes from essentially batch, nonintegrated systems to large-scale commercial facilities in one step is not yet feasible. The degree of scale up proposed by UAH is less ambitious than that by TVA.
- The acid hydrolysis stage of the TVA process has been tested in a 2 ton/day batch reactor. Fermentation of the hydrolyzate from the first stage has not yet been demonstrated. (Metals in the MSW may prove toxic to the yeast, among other considerations.)
- TVA laboratory and pilot plant data do not support the assumptions made in the process design for sugar yields from hydrolysis of WDF (waste derived feedstock).
- The MSW sterilization stage of the UAH process has been tested in a 2 ton/day batch reactor (600 lb/batch). Experimental data for the UAH process are incomplete.
- The experimental performance of the UAH reverse osmosis unit does not support the performance assumption made in the process design.
- Economic analyses prepared by both developers appear to rely on overly optimistic estimates for capital costs and revenues from byproducts and recycling. Increased paper recycling efforts will reduce the quantity of hydrolyzable cellulose in the waste stream, and increased glass, plastic, and metal recycling efforts will reduce the byproduct credits. On the other hand, increasing landfill tipping fees will favorably affect the economics of MSW processing.
- The processes need to be demonstrated using an MSW feedstock representative of current, "real life" conditions.

The facilities in question are essentially waste processing projects, the primary objective of which is the reduction of the waste stream that would otherwise ultimately be landfilled. Their development should be approached from that perspective. They should not be considered strictly ethanol (or energy) producing plants, although both liquid and solid fuels as well as electricity can be generated as byproducts.

Researchers conducted preliminary economic evaluations of 400 ton/day MSW processing plants based on either the dilute acid or enzymatic hydrolysis processes. An important difference exists between this analysis and those performed by TVA and UAH. This analysis assumes U.S. Government ownership of the facility, making such factors as depreciation, taxes, and insurance irrelevant to the analysis. The



analysis also assumes that the facilities operate for a 20-year lifetime at a rate of 24 hours/day, 330 days/year. The analysis uses a discount rate of 4.6 percent, which is consistent with DOE/DOD guidelines for analysis of federal energy management programs. The results of the economic evaluations in terms of simple payback period, net present value, and internal rate of return are:

1. The total capital investment for the base case dilute acid hydrolysis process was \$51.1 million; that for the base case enzymatic hydrolysis process was \$63.2 million.
2. The operating costs were determined to be \$8.0 million and \$7.3 million/year, respectively.
3. The tipping fee for the base case was set at \$45/ton (the U.S. average).
4. The simple payback periods for the dilute acid hydrolysis and enzymatic hydrolysis processes were 12.15 and 13.20 years, respectively.
5. The net present values were, respectively, positive \$3.1 million and negative \$1.5 million.
6. The internal rates of return were 5.3 percent and 4.3 percent, respectively.

Based on this analysis, the TVA process is marginally preferable. The anticipation of escalating tipping fees would be the primary economic justification for constructing these facilities.

In general, about 75 percent of the revenue for these plants is derived from tipping fees and sale of recyclable materials. Ethanol production accounts for only 18 to 19 percent of total facility revenues.

In terms of waste reduction/minimization, the UAH facility is somewhat more efficient than the TVA facility, with a net reduction of 62 percent compared with the TVA's 43 percent. As a result of the heavy reliance on tipping fees for income, both processes display a very high sensitivity to market changes in tipping-fee amounts.

The two processes contrast in that the TVA process experiences significantly larger landfill requirements, while the UAH process experiences much greater expenditures for electricity. Neither facility benefits from the sale of electricity as originally projected, although both do offset a portion of their own electric requirements by self-generation with residual solid fuel.

Important to the economics of MSW to fuel ethanol plants was the long-term stability of the MSW supply and the consistency of the MSW composition fed to the process. It is likely that the trend toward increased recycling efforts by local communities will reduce the quantity of recoverable and recyclable materials in the MSW stream, which will negatively impact the tipping fees required. Similarly, the quantity of cellulosic materials in the MSW stream will also likely decline, requiring that MSW be collected from a larger radius—at increasing transportation costs—to fill the existing plant capacity. Other factors, such as the suitability of the residual fraction of MSW remaining after processing for direct combustion for power generation (e.g., the ash fouling characteristics), and the environmental impact (HCl and CO<sub>2</sub> emissions), of MSW processing plants with fuel ethanol production capability, must be addressed by applying existing technologies.

## Recommendations

At their present state of individual development, neither the TVA nor UAH process is ready for scale-up to an economically viable, commercial-scale application. Both processes are advanced enough, however, to serve as a combined base for detailed engineering and technology development studies leading to design, construction, and operation of a commercial demonstration plant. It is recommended that:

- Combustion tests be conducted with the residue generated from enzymatic as well as acid hydrolysis to evaluate the level of acid-gas emissions that will occur
- Detailed environmental impact assessments be prepared prior to any further larger scale process development to identify areas of environmental sensitivity
- Integrated, continuous pilot plant scale testing of the processes be investigated in adequate detail at an equivalent, unprocessed MSW feed rate to the plant of about 20 tons/day. This would include MSW classification (to produce a waste derived feedstock) followed by pretreatment (prehydrolysis), hydrolysis (acid and enzymatic), acid neutralization/enzyme recovery, glucose/xylose fermentation, and distillation.
- Distillation of the fermentation beer, which will generate stillage bottoms, be analyzed for suitability as animal feedstock or other use. This investigation, which could be conducted during a focused 2-year period, will ensure the logical and economic progression of both processes to commercialization. The economics of the process should then be reevaluated in light of the more closely defined process performance.

## METRIC CONVERSION TABLE

1 ft	=	0.305 m
1 sq ft	=	0.093 m <sup>2</sup>
1 cu ft	=	0.028 m <sup>3</sup>
1 lb	=	0.453 kg
1 gal	=	3.78 l
1 psi	=	6.89 kPa

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**APPENDIX A: TVA and UAH Trip Report by Dr. Cavit Akin**



**TO: Gary W. Schanche**

**FROM: Cavit Akin**

**DATE: April 25, 1991**

**SUBJECT: PAPER WASTE TO ETHANOL PROCESSES OF TVA AND UAH**

The objective of this note is to summarize my first impressions and observations of the paper waste to ethanol processes of TVA and UAH while they are fresh in my mind. My comments are based on the oral presentations of TVA and UAH representatives, and the laboratory and pilot plant tours on April 23 and 24, 1991.

**OBJECTIVE OF TVA AND UAH PROCESSES:**

To produce ethanol for use as liquid transportation fuel from paper waste.

**OBJECTIVE OF OUR WORK:**

To investigate both the acid and enzymatic hydrolysis processes, their current state of development, and develop a preliminary assessment as to the potential for success of each project.

**PRELIMINARY ASSESSMENT:**

o At their present state of development, TVA and UAH processes are not ready for scale-up to economically viable commercial applications.

o Both of these processes are advanced enough to serve as a combined base for detailed engineering and technology development studies leading to design, construction and operation of a commercial demonstration plant.

**CHARACTERISTICS OF THE TVA AND UAH PROCESSES**

o Both processes initially aimed to convert woody and agricultural biomass to liquid fuels during the high oil prices period and moved into conversion of municipal solid waste as the source of cellulose when the environmental issues became the center of concern.

o TVA process relies on acid hydrolysis of cellulose for the generation of fermentable sugars. UAH process uses enzymatic hydrolysis.

o UAH process was tested in pilot plant scale in full, TVA process has been tested up to fermentation stage. TVA fermentation pilot plant is ready and tested with fermentation broth prepared with purchased glucose and plans are underway to conduct tests with the acid hydrolysate.

o Both processes borrowed or adapted equipment and technologies from a variety of industrial practices ranging from paper production to meat rendering. Both lack an integrated engineering approach. The types of equipment and their operational layout need significant modifications.

o Economic analysis of both processes appear to have relied on some optimistic energy and mass balances; raw material composition which may not hold true in the near future; revenue streams with limited localized demand; and an average tipping fee which may be too high or too low according to the location of the plant. Assumptions such as "steam is free" and "energy credit for residue burning (not tested)" result in unreliable (low) cost estimates.

o In the TVA process the time, temperature, pressure conditions that were considered optimum in the bench-scale studies do not appear to be reproduced in the pilot-scale operations.

o In the UAH process operating conditions appear to be determined empirically to fit the equipment and energy sources available at the time of experimentation.

#### WHAT NEEDS TO BE DONE:

o "Solving the environmental issues relating to MSW and other paper containing wastes" should be considered as the main driving force for the commercialization of TVA and UAH processes. Both of these processes should be reassessed on the basis of this environmental driving force. Liquid fuel production, if economical, can be considered as an added benefit.

o Each step of the TVA and UAH processes need be re-evaluated in respect to reaction kinetics and mechanism, material and energy balances, achievable concentrations of fermentable sugars and ethanol, new or added environmental burden, capital and operational costs, mechanical reliability, operational ease, and competing approaches.

o Various groups in the United States are working on similar projects. If a major capital investment is envisioned, a fair assessment of the other processes (e.g. SERI process) would be prudent. Also needed are the opinion and input of companies currently producing grain alcohol, and companies involved in MSW collection and treatment.

#### **SUGGESTIONS:**

o TVA has the needed personnel and equipment to conduct an expanded pilot plant study over and beyond what was accomplished so far. This study should aim to provide a sound engineering basis for the design and operation of a commercial-scale plant. The scope of the study should include the integration of advantageous elements of the UAH process and processes developed by other groups, cost cutting, process simplification, and successful demonstration of the fully integrated process in the pilot plant.

o In dealing with MSW conversion, a stepwise approach to commercialization should be considered. For example a process that can screen and sort MSW to recyclable components in a cost effective manner may be adequate to solve some local MSW landfill problems. Also using the low moisture solid organic components of MSW as a fuel in power generators may turn out to be more cost effective than production of ethanol as a liquid fuel.

o The changes in waste generation patterns, such as waste reduction at source and curbside recycling, are expected to alter the composition of MSW in the near future. Process schemes based on rigid expectations of the MSW composition may become obsolete within 5 to 10 years. MSW processing plants need be flexible to deal with a variable MSW composition. Their capital demands should be low.

o In the enzymatic hydrolysis of cellulose and ethanol fermentation steps, progress in molecular genetics should be considered for the development of more efficient processes.

o In acid hydrolysis of cellulose, application of high temperature short time techniques should be considered in order to achieve high conversion rates, high fermentable sugars yield, high sugar concentration and low concentration of undesirable byproducts. (I believe there are techniques that would eliminate the need for a high-pressure vessel, and improve the residue separation and collection steps).

o A results oriented two year intensive pilot plant study plan should be developed and implemented at TVA with funds contributed by TVA and other interested groups. The general project management and coordination should be provided by the U.S. Corps of Engineers.

**APPENDIX B: Revised Material Balances for TVA and UAH Processes**

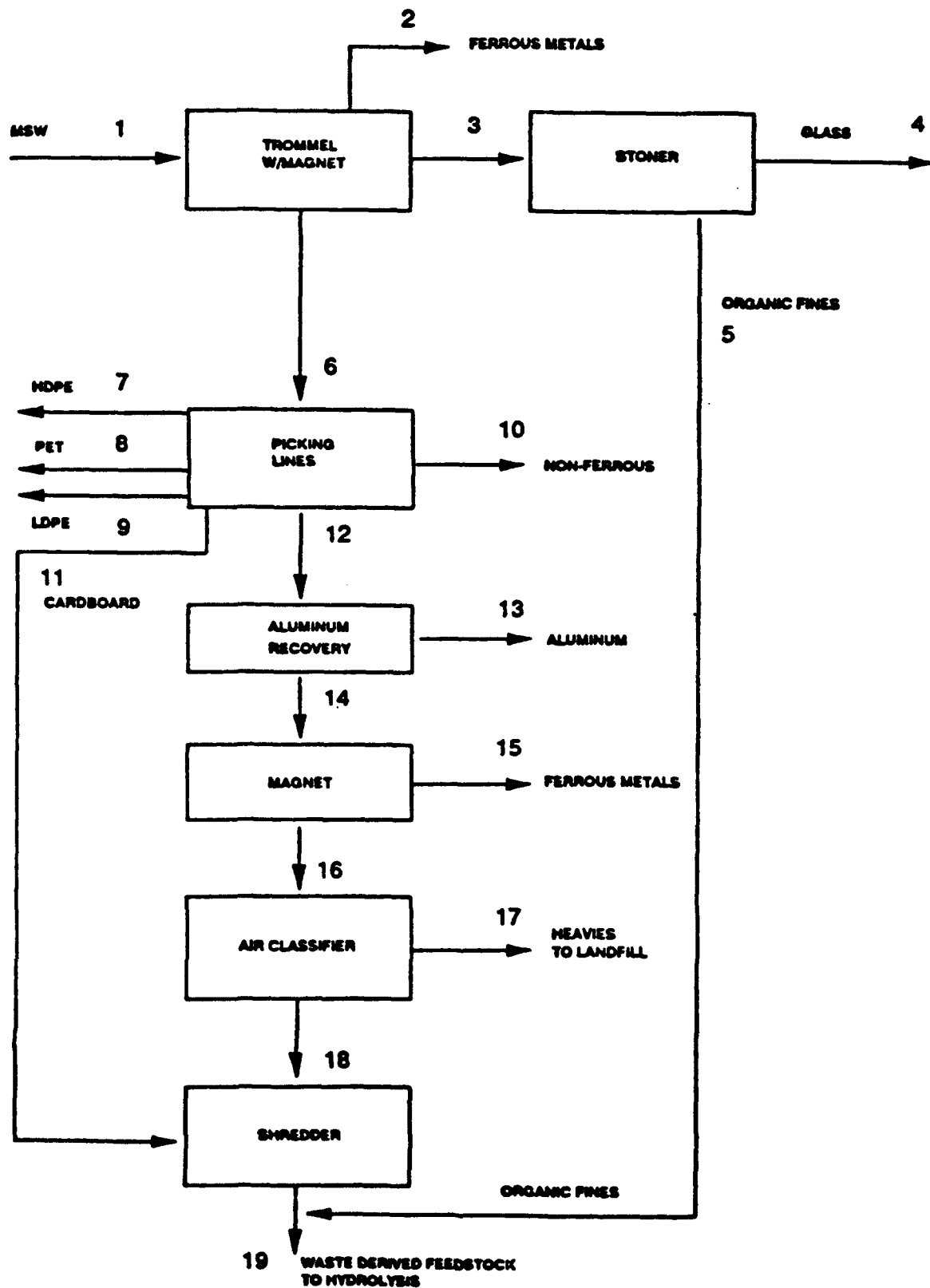


Figure 6. Flow diagram of front-end classification system.

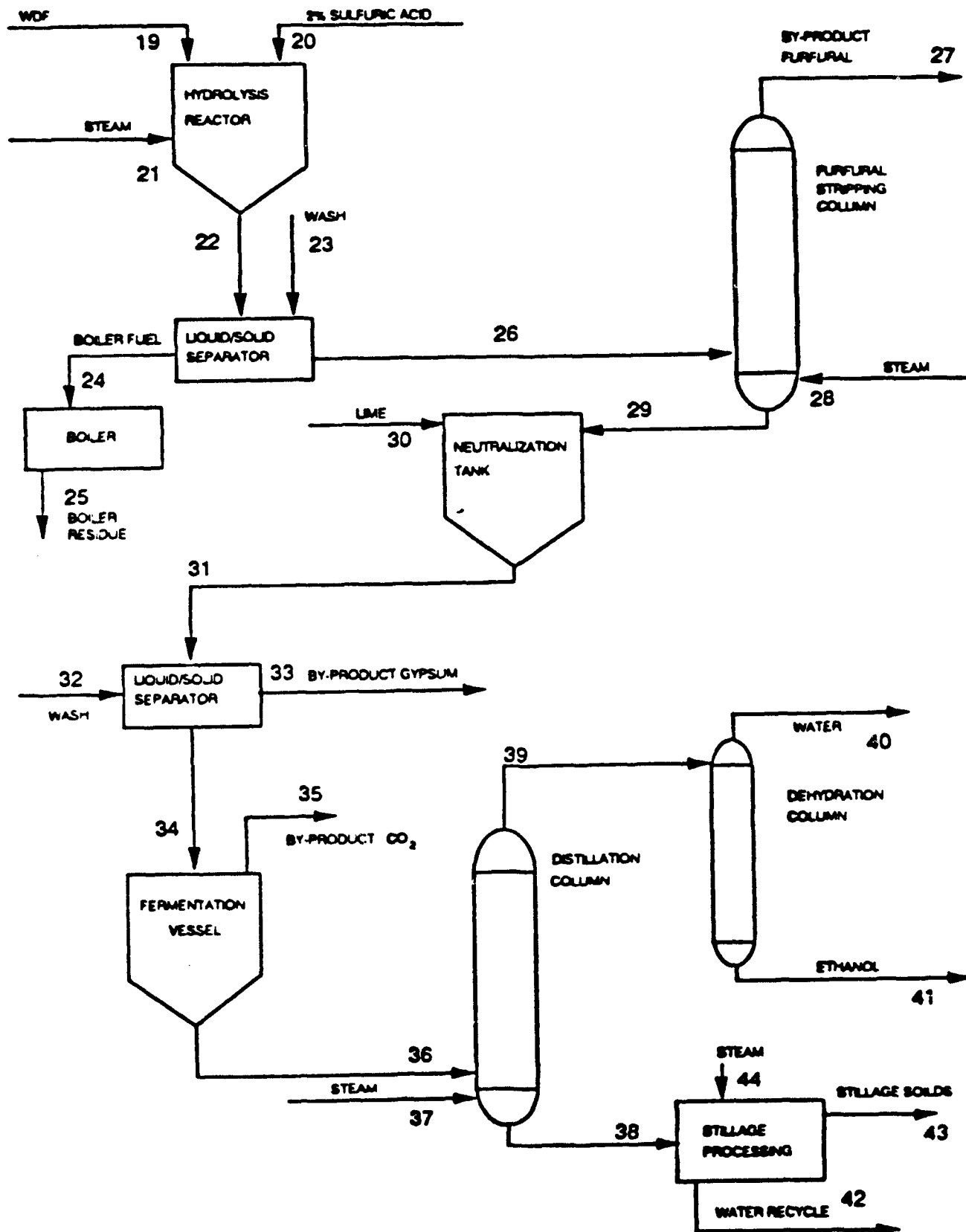


Figure 7. Flow diagram of WDF hydrolysis processing system.

Table 16. Mass balance for the 400-ton-a-day MSW processing facility.

STREAM NUMBER			1	2	3	4	5	6	7
STREAM COMPONENT	QUANTITY OF MSW (TPD)	MSW COMPOSITION (PERCENT)	MSW TO TROMMEL (TPD)	FERROUS METALS FROM TROMMEL	FINES TO STONER	GLASS FROM STONER	ORGANIC FINES TO HYDROLYSIS	WDF TO PICKING LINES	HDPE TO BALER
MSW	400.00								
PAPER & PAPERBOARD		30.98	123.9 <sup>2</sup>					123.90	
GLASS		9.90	39.60		39.60	33.66	5.94		
ALUMINUM		1.40	5.60					5.60	
FERROUS METALS/ TIN/STEEL		8.85	35.40	26.55				8.85	
NON-FERROUS METALS		0.25	1.00					<del>0.05</del> 1.00	
HDPE		2.70	10.80					10.80	10.26
LDPE		3.20	12.80					12.80	
PET		1.40	5.60					5.60	
OTHER PLASTICS		2.80	11.20					11.20	
RUBBER & LEATHER		2.81	11.2 <sup>4</sup>					11.22	
TEXTILES		1.92	7.68					7.68	
FOOD WASTES		3.68	14.7 <sup>2</sup>		14.70	0.74	13.97		
WOOD		3.60	14.40		14.40	0.72	13.68		
YARD WASTES		2.55	10.20		10.20	0.51	9.69		
MISC. INORGANICS		2.10	8.40		8.40	0.42	7.98		
WATER		21.88	87.5 <sup>2</sup>		10.26	0.91	17.34	34.19	
TOTAL	400.00	100.00	400.00	26.55	105.56	36.96	<del>82.46</del> 66.60	<del>237.89</del> 232.84	10.26

Table 16. continued.

STREAM NUMBER	8	9	10	11	12	13	14	15	16	17	18	19
STREAM COMPONENT	PET TO BALER	LDPE TO LANDFILL	NON-FERROUS TO STORAGE	PAPER TO SHREDDER	PICKING LINES TO ALUMINUM SEPARATOR	ALUMINUM TO BALER	WDF TO MAGNET	METALS FROM MAGNET	MAGNET RESIDUE TO AIR	HEAVIES TO LANDFILL	LIGHTS TO SHREDDER	WDF TO HYDROLYSIS
MSW	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
PAPER & PAPERBOARD	-----	-----	-----	74.34	49.56	-----	49.56	-----	49.56	2.48	47.08	121.42
GLASS	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	5.94
ALUMINUM	-----	-----	-----	-----	<del>5.60</del> 5.04	5.04	0.56	-----	0.56	-----	0.56	0.56
FERROUS METALS/ TIN/STEEL	-----	-----	-----	-----	8.85	-----	8.85	7.08	1.77	1.42	<del>0.35</del>	0.35
NON-FERROUS METALS	-----	-----	0.95	-----	0.05	-----	0.05	0.05	-----	-----	-----	-----
MDPE (milk)	-----	-----	-----	-----	0.54	-----	0.54	-----	0.54	-----	0.54	0.54
LDPE	-----	12.80	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
PET (soda)	5.32	-----	-----	-----	0.25	-----	0.25	-----	0.25	-----	0.25	0.28
OTHER PLASTICS	-----	-----	-----	-----	11.20	-----	11.20	-----	11.20	10.64	<del>0.56</del>	0.56
RUBBER & LEATHER	-----	-----	-----	-----	11.22	-----	11.22	-----	11.22	9.35	1.87	1.87
TEXTILES	-----	-----	-----	-----	7.68	-----	7.68	-----	7.68	5.44	2.24	2.24
FOOD WASTES	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	13.97
WOOD	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	13.68
YARD WASTES	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	9.69
MISC. INORGANICS	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	7.98
WATER	-----	-----	-----	18.59	15.61	-----	15.61	-----	15.61	3.11	12.50	83.48
TOTAL	5.32	12.80	0.95	92.93	<del>109.21</del> 110.57	5.04	<del>104.71</del> 105.45	7.13	<del>99.83</del> 98.42	32.43	<del>67.84</del> 65.93	262.51



Table 16. continued.

STREAM NUMBER				20	21	22	23	24	25	26	27
STREAM COMPONENT	QUANTITY OF DRY WDF (TPD)	PERCENTS OF WDF (WET)	BIOMASS FEEDSTOCK	SULFURIC ACID SOLUTION	STEAM	ACID/ BIOMASS TO PRESS	WASH WATER	SOLIDS TO BOILER	BOILER RESIDUE TO LANDFILL	HYDROLYZATE TO FURFURAL STRIPPING	FURFURAL AND ACETIC ACID
WDF	179.09	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
HEMI- CELLULOSE	-----	3.75	9.85	-----	-----	1.97	-----	0.63	-----	-----	-----
CELLULOSE	-----	37.79	99.21	-----	-----	61.51	-----	61.51	-----	-----	-----
LIGNIN	-----	10.57	27.76	-----	-----	27.76	-----	27.76	-----	-----	-----
OTHER	-----	7.50	19.70	-----	-----	19.70	-----	19.70	1.97	-----	-----
ASH	-----	8.59	22.56	-----	-----	22.56	-----	22.56	22.56	-----	-----
XYLOSE	-----	-----	-----	-----	-----	4.92	-----	0.15	-----	4.78	-----
GLUCOSE	-----	-----	-----	-----	-----	37.70	-----	1.13	-----	36.57	-----
FURFURAL	-----	-----	-----	-----	-----	2.95	-----	0.09	-----	2.87	2.87
SULFURIC ACID	-----	-----	-----	10.75	-----	10.75	-----	0.32	-----	10.42	-----
WATER	-----	31.79	83.48	321.07	121.97	526.52	137.84	133.03	-----	531.33	-----
ACETIC ACID	-----	-----	-----	-----	-----	-----	-----	-----	-----	1.34	1.34
CARBON DIOXIDE	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
LIME	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
GYPSON	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
ETHANOL	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
TOTAL	179.09	100.00	262.56	331.82	121.97	716.35	137.84	266.89	24.53	587.31	4.21

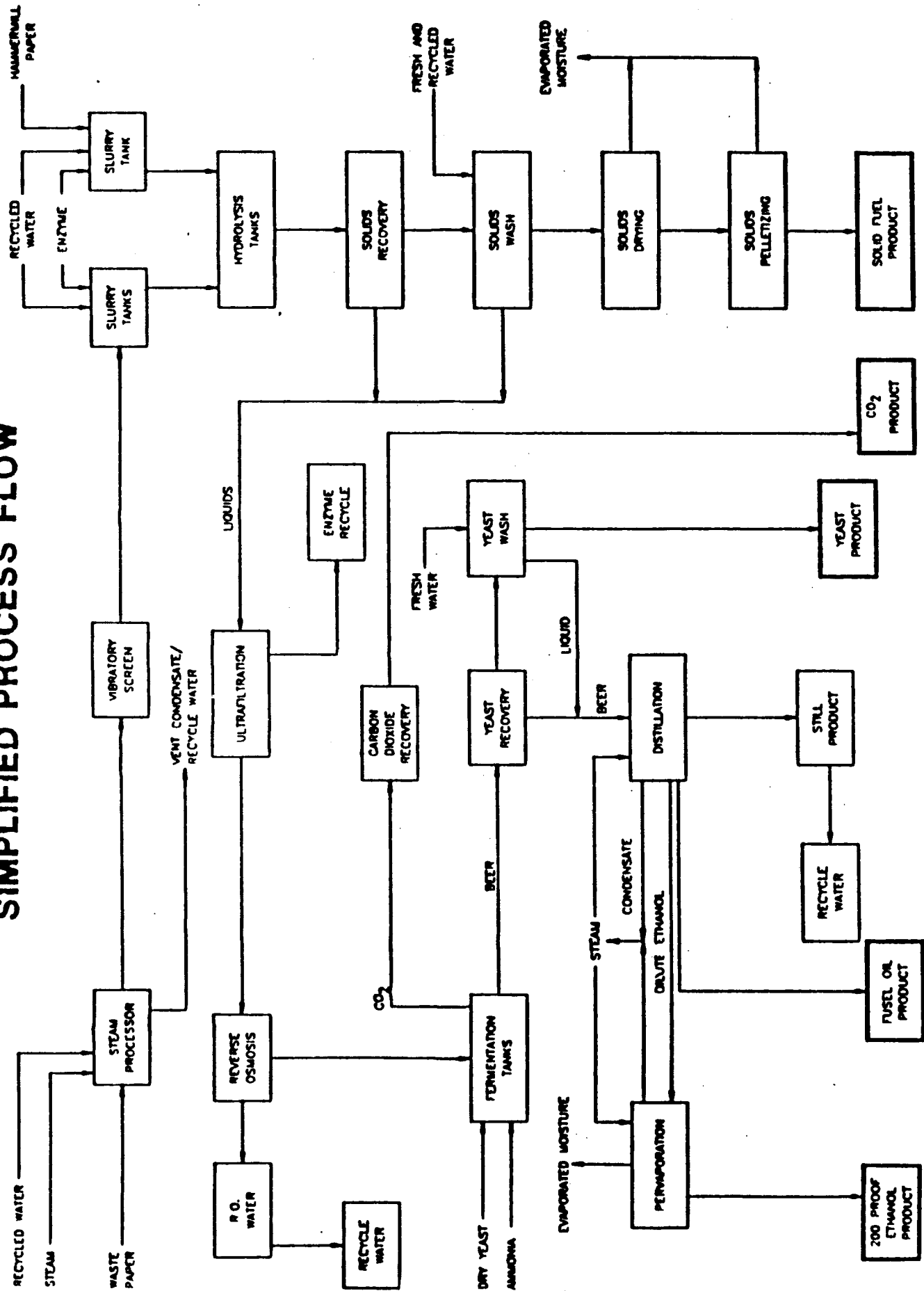
Table 16. continued.

STREAM NUMBER	28	29	30	31	32	33	34	35	36
STREAM COMPONENT	STEAM	HYDROLYZATE TO NEUTRAL- IZATION	LIME	HYDROLYZATE GYPSUM SOLUTION	WASH WATER	GYPSUM TO DISPOSAL	HYDROL. TO FER- MENTATION	CARBON DIOXIDE	BEER TO DISTILLATION
WDF									
HEMI- CELLULOSE									
CELLULOSE									
LIGNIN									
OTHER									
ASH									
XYLOSE		4.78		4.78		0.14	4.63		0.46
GLUCOSE		36.57		36.57		1.10	35.47		3.55
FURFURAL									
SULFURIC ACID		10.42							
WATER	48.42	531.33		531.33	57.16	45.58	542.91		542.91
ACETIC ACID									
CARBON DIOXIDE								17.69	
LIME			7.87						
GYPSUM				18.29		18.29			
ETHANOL									18.41
TOTAL	48.42	583.10	7.87	590.97	57.16	65.11	583.01	17.69	565.33

Table 16. continued.

STREAM NUMBER	37	38	39	40	41	42	43	44
STREAM COMPONENT	STEAM	STILLAGE TO EVAPORATOR	ETHANOL TO DENYDRATION	WATER FROM DENYDRA.	ETHANOL FROM DENYDRATION	MAKE-UP WATER TO STORAGE	EVAPORATOR STILLAGE TO BOILER	STEAM
WDF	-----	-----	-----	-----	-----	-----	-----	-----
HEMI- CELLULOSE	-----	-----	-----	-----	-----	-----	-----	-----
CELLULOSE	-----	-----	-----	-----	-----	-----	-----	-----
LIGNIN	-----	-----	-----	-----	-----	-----	-----	-----
OTHER	-----	-----	-----	-----	-----	-----	-----	-----
ASH	-----	-----	-----	-----	-----	-----	-----	-----
XYLOSE	-----	0.46	-----	-----	-----	-----	0.46	-----
GLUCOSE	-----	3.55	-----	-----	-----	-----	3.55	-----
FURFURAL	-----	-----	-----	-----	-----	-----	-----	-----
SULFURIC ACID	-----	-----	-----	-----	-----	-----	-----	-----
WATER	59.40	541.97	0.94	0.94	-----	539.29	2.67	180.64
ACETIC ACID	-----	-----	-----	-----	-----	-----	-----	-----
CARBON DIOXIDE	-----	-----	-----	-----	-----	-----	-----	-----
LIME	-----	-----	-----	-----	-----	-----	-----	-----
GYP SUM	-----	-----	-----	-----	-----	-----	-----	-----
ETHANOL	-----	0.55	17.86	-----	17.86	0.55	-----	-----
TOTAL	59.40	546.53	18.80	0.94	17.86	539.84	6.68	180.64

# SIMPLIFIED PROCESS FLOW



**NET DAILY MATERIAL BALANCE**  
**(Based on 16 & 27 Ton/Day)**

<u>IN</u>		<u>16 TON</u>	<u>27 TON</u>	
Waste Paper	-	22,000	44,000	lbs/day
Hammermilled Paper	-	10,000	10,000	lbs/day
Steam	46,895 -	<del>41,935</del>	67,795 <del>78,385</del>	lbs/day
Water	-	28,177	47,216	lbs/day
Enzyme	-	152	257	lbs/day
Yeast	-	43	75	lbs/day
Ammonia	-	<u>871</u>	<u>1,508</u>	lbs/day
		<del>103,179</del>	<del>181,341</del>	lbs/day
		108,138	170,851	

<u>OUT</u>		<u>16 TON</u>	<u>27 TON</u>	
Water Recycled	-	<del>14,143</del>	<del>27,743</del>	<del>lbs/day</del>
Steam Condensate Recycled	46,895 -	<del>28,350</del>	67,795 <del>51,115</del>	lbs/day
Evaporated Moisture	-	26,324	44,422	lbs/day
Solid Fuel Product	-	20,728	35,051	lbs/day
CO <sub>2</sub> Product	-	6,334	10,687	lbs/day
Yeast Product (70% Moisture)	-	1,054	1,784	lbs/day
Fusel Oil Product	-	<del>557</del>	<del>573</del>	lbs/day
200-Proof Ethanol Product	-	<u>6,246</u>	<u>10,539</u>	lbs/day
		<del>103,179</del>	<del>181,341</del>	lbs/day
		108,138	170,851	

## STEAM CLASSIFICATION

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
1 Paper	- 22,000	44,000	lbs/day
2 Hotwater	- 19,800	39,600	lbs/day
3 Steam	- <u>20,900</u>	<u>41,800</u>	lbs/day
	62,700	125,400	lbs/day

<u>OUT</u>	<u>16 TON</u>	<u>27 TON</u>	
4 Processor Steam Vent	- 7,315	14,630	lbs/day
8 Cellulosics to Conversion	- <u>55,385</u>	<u>110,770</u>	lbs/day
	62,700	125,400	lbs/day

## 2 HOT WATER

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
4 Steam Vent Condensate	- 1,759	3,980	lbs/day
5 Still Bottoms	- <u>18,041</u>	<u>35,620</u>	lbs/day
	19,800	39,600	lbs/day

<u>OUT</u>			
2 Hot water to Steam Processor	19,800	39,600	lbs/day

## SOLID FUEL RECOVERY

<u>IN</u>		<u>16 TON</u>	<u>27 TON</u>	
16	Hydrolysis Slurry	- 1,219,817	2,064,661	lbs/day
21	Solids Wash Water	- <u>89,150</u>	<u>150,523</u>	lbs/day
	<i>STEAM</i>	<i>25995</i>	<i>?</i>	
		1,308,967	2,215,184	lbs/day

<u>OUT</u>		<u>16 TON</u>	<u>27 TON</u>	
25	Fuel Moisture Evaporated	- 25,995	43,867	lbs/day
26	Solid Fuel Product	- 20,728	35,051	lbs/day
29	Liquids to Ultrafilter	- <u>1,262,244</u>	<u>2,136,266</u>	lbs/day
	<i>CONDENSATE</i>	<i>25995</i>	<i>?</i>	
		1,308,967	2,215,184	lbs/day

## 21 SOLIDS WASH WATER

<u>IN</u>		<u>16 TON</u>	<u>27 TON</u>	
4	Processor Steam Vent	- 5,556	10,650	lbs/day
5	Still Bottoms	- 56,994	95,326	lbs/day
20	Fresh Water Make-up	- <u>26,600</u>	<u>44,547</u>	lbs/day
		89,150	150,523	lbs/day
<u>OUT</u>				
21	<i>WATER TO SOLIDS WASH</i>	<i>89,150</i>	<i>150,523</i>	<i>lb/day</i>

## CELLULOSICS TO HYDROLYSIS

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
8 Processed Cellulosics	- 55,385	110,770	lbs/day
11 Hammermilled Paper	- 10,000	10,000	lbs/day
12 Dilute Enzyme Solution	- <u>1,154,432</u>	<u>1,943,891</u>	lbs/day
	1,219,817	2,064,661	lbs/day

<u>OUT</u>	<u>16 TON</u>	<u>27 TON</u>	
16 Slurry to Hydrolysis	- 1,219,817	2,064,661	lbs/day

## 12 DILUTE ENZYME SOLUTION

*ENZYME RECYCLE STORAGE*

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
6 R. O. Water	- 1,089,521	1,833,662	lbs/day
14 U. F. Concentrate	- 64,759	109,972	lbs/day
15 Dry Make-up Enzyme	- <u>152</u>	<u>257</u>	lbs/day
	1,154,432	1,943,891	lbs/day

OUT

12 <i>ENZYME RECYCLE TO SLURRY TANKS</i>	1,154,432	1,943,891	lbs/day
--	-----------	-----------	---------



# **FERMENTATION AND ENZYME, SUGAR, AND YEAST RECOVERY**

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
29 Liquids to Ultrafiltration	- 1,262,244	2,136,266	lbs/day
35 Dry Yeast	- 44	75	lbs/day
36 Ammonia	- 871	1,508	lbs/day
20 Fresh Water	- <u>1,577</u>	<u>2,669</u>	lbs/day
	1,264,736	2,140,518	lbs/day
<u>OUT</u>	<u>16 TON</u>	<u>27 TON</u>	
14 UF Concentrate/ Enzyme Recycle	- 64,759	109,972	lbs/day
6 R. O. Water	- 1,110,422	1,875,462	lbs/day
37 Carbon Dioxide (CO <sub>2</sub> )	- 6,334	10,687	lbs/day
40 Yeast (70% Moisture)	- 1,054	1,784	lbs/day
42 Beer to Distillation	- <u>82,167</u>	<u>142,613</u>	lbs/day
	1,264,736	2,140,518	lbs/day

### ETHANOL RECOVERY

<u>IN</u>		<u>16 TON</u>	<u>27 TON</u>	
42 Beer to Distillation	-	82,167	142,613	lbs/day
3 Steam to Distillation	-	20,542	35,653	lbs/day
3 Steam to Pervaporation	-	<u>493</u>	<u>832</u>	lbs/day
		103,202	179,098	lbs/day

<u>OUT</u>		<u>16 TON</u>	<u>27 TON</u>	
<i>FUSOL 0.1</i>		<i>557</i>	<i>573</i>	
44 200-Proof Ethanol	-	6,246	10,598	lbs/day
5 Still Bottoms	<i>75035</i>	<del>75,592</del>	<del>131,519</del>	lbs/day
25 Evaporated Moisture	-	329	555	lbs/day
45 Dist. Reboiler Condensate	-	20,542	35,653	lbs/day
45 Pervap. Reboiler Condensate	-	<u>493</u>	<u>832</u>	lbs/day
		103,202	179,098	lbs/day

## **APPENDIX C: Communications with TVA and UAH**

Questions concerning the material balances, processing steps, or other areas presented in the two major subject documents were submitted to the process developers in letter form. A copy of each letter was also sent to the CERL project manager concurrently. In the interests of time, this approach was used rather than first direct questions to CERL to be forwarded to the process developers.

Copies of the letters sent are attached. The responses of the developers are also attached.

# GT

adquarters

September 6, 1991

Dr. J. Wayne Barrier  
Program Manager  
Biotechnical Research Department  
Tennessee Valley Authority  
CEB 5C-M  
P. O. Box 1010  
Muscle Shoals, Alabama 35660-1010

Dear Dr. Barrier:

You may or may not be aware that the U.S. Army Construction Engineering Research Laboratory (CERL) has contracted the Institute of Gas Technology (IGT, Chicago) to conduct a process and economic evaluation of the Preliminary Feasibility Study of the Recovery and Production of Recyclables, Chemicals, and Electricity From Municipal Solid Waste for Southern Indiana, which was prepared by TVA.

IGT has made a preliminary review of the material included in the above-referenced feasibility study and has prepared a list of questions and comments that arose during the evaluation. Most of the questions or comments relate to the material balance or to processing steps or conditions that have not been explicitly identified. The list of questions and comments is attached.

It would greatly facilitate IGT's completion of the process review if you, or your designee, could answer the questions and provide the needed information. As you can imagine, time is of the essence.

If the information could be faxed to IGT (facsimile no. 312-567-5209) by September 18 or earlier, it would be greatly appreciated. Additional questions may be faxed to you as they arise.

Thank you in advance for your cooperation and prompt attention to this request. Please call me at (312)-567-3730, or Dr. Goyal at (312)-567-5759 if you have questions about any of the attached comments.

Sincerely,



Michael C. Mensinger  
Manager  
Process Development

cc: Dr. Mike Lin, CERL  
C. Blazek  
A. Goyal

Institute of Gas Technology

Headquarters  
3424 South State Street  
Chicago, Illinois 60616  
(312) 567-3650  
TELEX 25-6185  
FAX (312) 567-5209

C-2

Energy Development Center  
4201 West 36th Street  
Chicago, Illinois 60632  
(312) 890-7000  
FAX (312) 890-6400

Washington Office  
1825 K Street, N.W.  
Washington, D.C. 20006  
(202) 785-3511  
FAX (202) 223-4111

Questions/Comments concerning the TVA Preliminary Feasibility Study of the Recovery and Production of Recyclables, Chemicals, and Electricity From Municipal Solid Waste for Southern Indiana.

Please refer to the enclosed material balance and process flow diagram for specific questions/comments.

1. Some material balances are violated. A copy of the flow sheet showing the corrected values is attached. Corrected material balance tables are also included.
2. There are species missing from several intermediate streams. However, all species are indicated in the WDF (Stream 19) except water content. A balance around the "Shredder" shows that the amount of water should be 48.43 tons instead of 83.48 tons. (This amount has been adjusted based on the WDF water composition given in Table 5, page 18).
3. The bases of steam feeds to Furfural Stripping Column (No. 28), Distillation Column (No. 37), and Stillage Processing (No. 44) are not known. These steam flows do not show up in Product streams 27 and 29, 38 and 39, or 42 and 43, respectively. Is the purpose of this steam to provide heat only, and in that case condensate leaving the column should be shown to reflect material balance.
4. The basis of the amount of wash water in streams 23 and 32 are not known.
5. The sulfuric acid in Stream 20 is 3.24 wt %. The specified concentration is 2 wt %. Is the specified concentration wt % or volume %.
6. The material balance table shows that about 68% of the hemicellulose entering the "Liquid/Solid Separator" stage via Stream 22 is converted to acetic acid. Is acetic acid really formed in this separator or is it actually formed in the "Hydrolysis Reactor?"
7. An air feed to the "Boiler" and flue gases leaving the "Boiler" should be shown.
8. The exponent of the quantity "Energy Required (Btu/hr)" in Table 14 should be positive instead of negative.

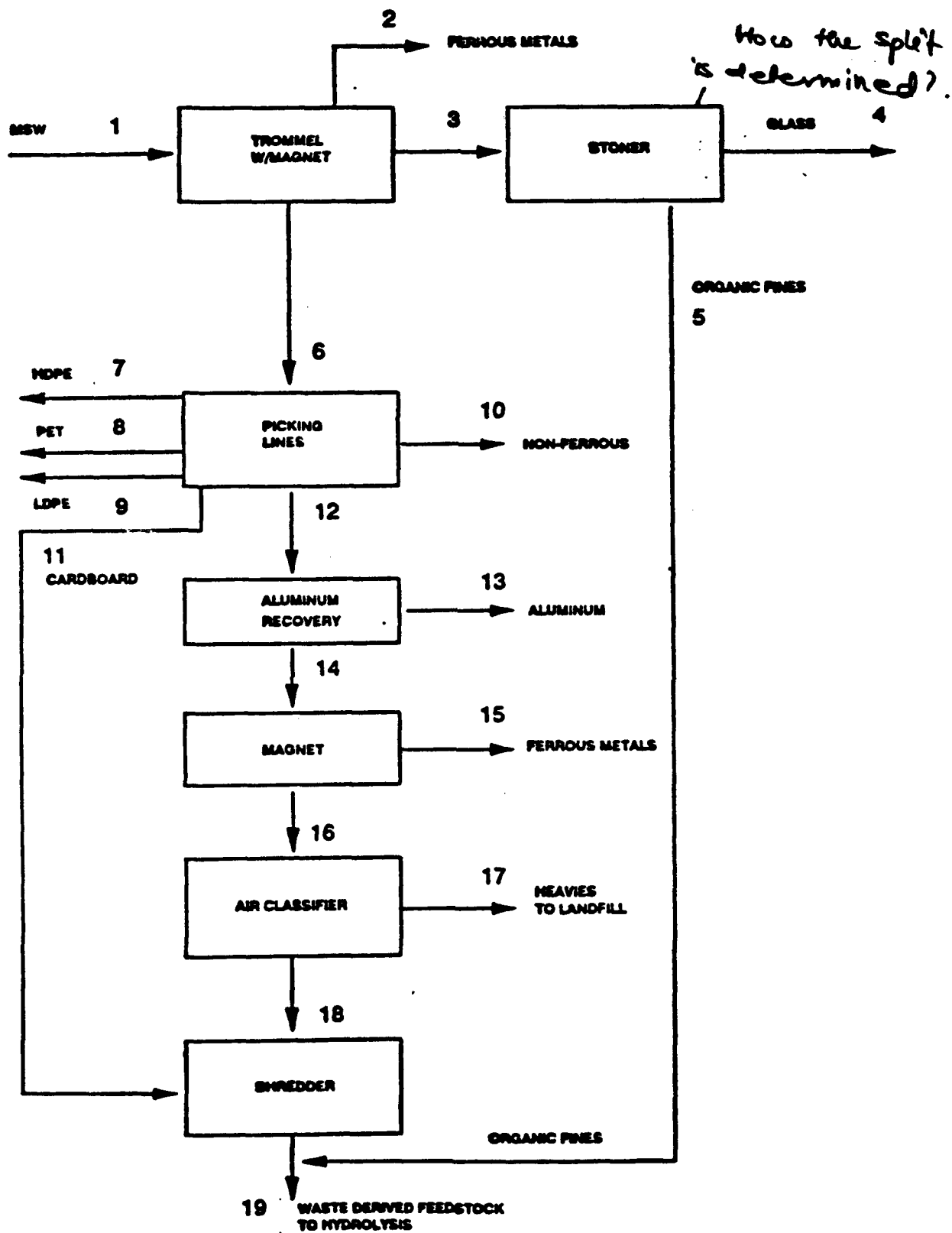


Figure 6. Flow diagram of front-end classification system.

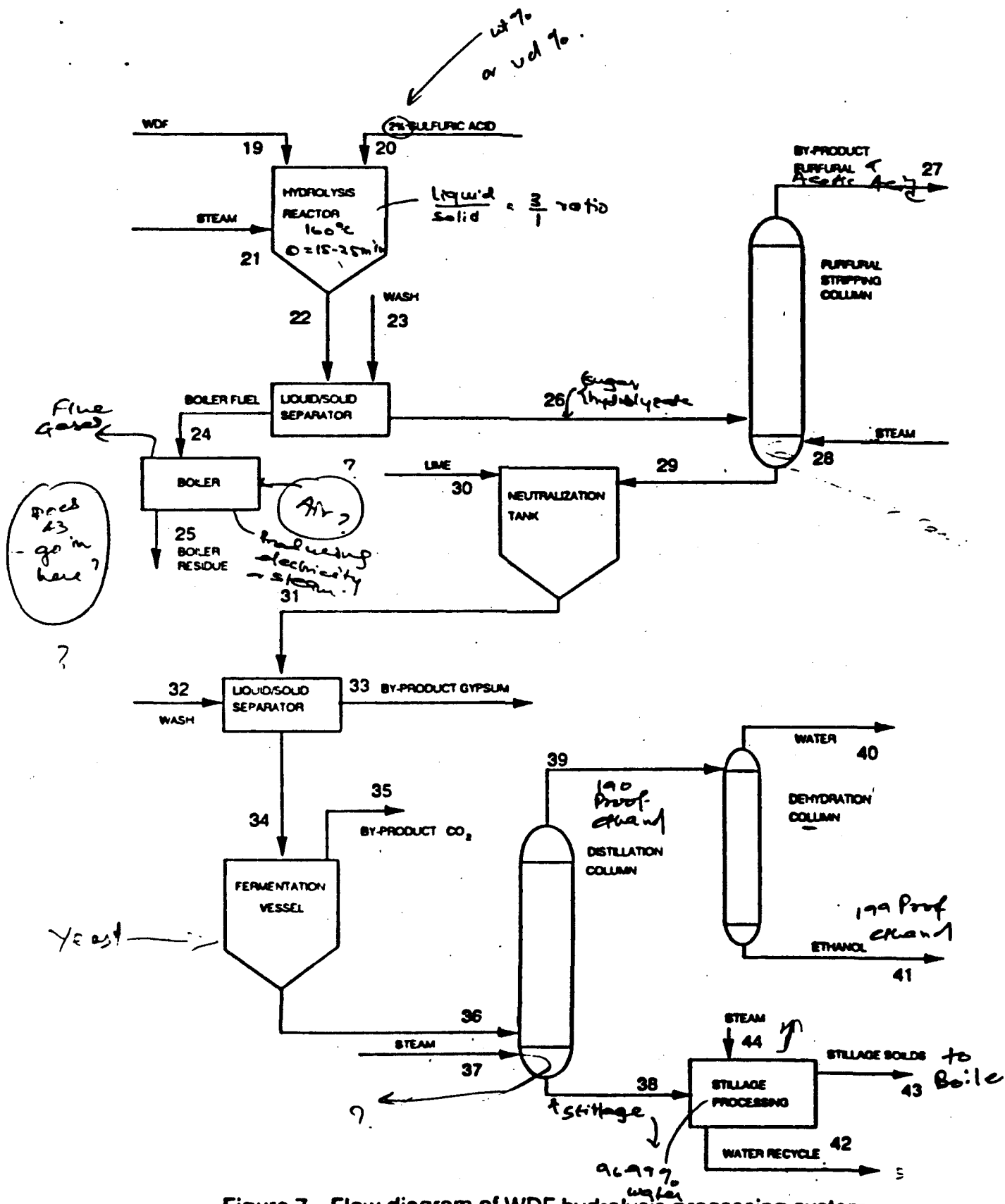


Figure 7. Flow diagram of WDF hydrolysis processing system.

$$1 = 2 + 3 + 6 \leftarrow ?$$

$$3 = 4 + 5 \leftarrow ?$$

$$4 = 7 + 8 + 9 + 10 + 11 + 12$$

Table 16. Mass balance for the 400-ton-a-day MSW processing facility.

STREAM NUMBER	✓	✓	1	2	3	4	5	6	7
STREAM COMPONENT	QUANTITY OF MSW (TPD)	MSW COMPOSITION (PERCENT)	MSW TO TROMMEL (TPD)	FERROUS METALS FROM TROMMEL	FINES TO STONER	GLASS FROM STONER	ORGANIC FINES TO HYDROLYSIS	WDF TO PICKING LINES	NOPE TO BALER
MSW	400.00								
PAPER & PAPERBOARD		✓ 30.98	✓ 123.97 <sub>2</sub>					✓ 123.90	
GLASS		✓ 9.90	✓ 39.60		✓ 39.60	✓ 33.66	5.94		
✓ ALUMINUM		✓ 1.40	✓ 5.60					✓ 5.60	✓
✓ FERROUS METALS/TIN/STEEL		✓ 8.85	✓ 35.40	✓ 26.55				✓ 8.85	✓
NON-FERROUS METALS		✓ 0.25	✓ 1.00					1.00 / 0.25	
NOPE <i>Plastics</i>		✓ 2.70	✓ 10.80					✓ 10.80	10.26
LDPE <i>Plastics</i>		✓ 3.20	✓ 12.80					✓ 12.80	
PET <i>Plastics</i>		✓ 1.40	✓ 5.60					✓ 5.60	
OTHER PLASTICS		✓ 2.80	✓ 11.20					✓ 11.20	
RUBBER & LEATHER		✓ 2.81	11.27 <sub>4</sub>					✓ 11.22	
TEXTILES		✓ 1.92	✓ 7.68					✓ 7.68	
FOOD WASTES		✓ 3.68	14.77 <sub>2</sub>		✓ 14.70	0.74	✓ 13.97		
WOOD		✓ 3.60	✓ 14.40		✓ 14.40	✓ 0.72	✓ 13.68		
YARD WASTES		✓ 2.55	✓ 10.20		✓ 10.20	✓ 0.51	✓ 9.69		
MISC. INORGANICS		✓ 2.10	✓ 8.40		8.40	0.42	7.98		
WATER		✓ 21.88	✓ 87.57 <sub>2</sub>		18.26	0.91	17.34	34.19	
TOTAL	✓ 400.00	✓ 100.00	✓ 400.00	✓ 26.55	105.56	34.96	62.55	234.89	10.26 ✓
									234.84



Table 16. continued.

$12 = 13 + 14$   
 $14 = 15 + 16$   
 $16 = 17 + 18$   
 $19 = 11 + 5 + 18 \leftarrow \text{water imbalance}$

STREAM NUMBER	8	9	10	11	12	13	14	15	16	17	18	19
STREAM COMPONENT	PET TO BALER	LDPE TO LANDFILL	NON-FERROUS PAPER TO STORAGE	PAPER SHREDDER	PICKING LINES TO ALUMINUM BALER	ALUMINUM TO MAGNET	MDF TO MAGNET	METALS FROM MAGNET	MAGNET RESIDUE TO AIR	HEAVIES TO LANDFILL	LIGHTS TO SHREDDER	MDF TO HYDROLYSIS
MSW	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
PAPER & PAPERBOARD	-----	-----	-----	74.34	49.56	-----	49.56	-----	49.56	2.48	47.08	121.42
GLASS	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	✓5.94
ALUMINUM	-----	-----	-----	5.60	5.04	0.56	-----	0.56	-----	0.56	0.56	0.56
FERROUS METALS/TIN/STEEL	-----	-----	-----	8.85	-----	8.85	7.08	1.77	1.42	0.35	0.35	0.35
NON-FERROUS METALS	-----	-----	0.95	-----	0.05	-----	0.05	0.05	-----	-----	-----	-----
MDPE (milk)	-----	-----	-----	0.54	0.54	0.54	0.54	0.54	0.54	0.54	0.54	0.54
LDPE	-----	12.80	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
PET (soda)	5.32	-----	-----	0.28	0.28	0.28	0.28	0.28	0.28	0.28	0.28	0.28
OTHER PLASTICS	-----	-----	-----	-----	11.20	-----	11.20	-----	11.20	10.64	0.56	0.56
RUBBER & LEATHER	-----	-----	-----	-----	11.22	-----	11.22	-----	11.22	9.35	1.87	1.87
TEXTILES	-----	-----	-----	-----	7.68	-----	7.68	-----	7.68	5.44	2.24	2.24
FOOD WASTES	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	13.97
WOOD	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	13.68
YARD WASTES	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	9.69
MISC. INORGANICS	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	7.98
WATER	-----	-----	-----	18.59	15.61	-----	15.61	-----	15.61	3.11	12.50	83.48
TOTAL	5.32	✓12.80	✓0.95	✓92.93	✓104.77	5.04	✓104.73	7.13	✓95.83	32.43	✓61.26	✓242.57
				110.59	34	105.55	98.42					MP = 179.1

Table 16. continued.

STREAM NUMBER	19	20	21	22	23	24	25	26	27		
STREAM COMPONENT	QUANTITY OF DRY MAT (TPO)	PERCENTS OF MAT (WET)	BIO MASS FEEDSTOCK SOLUTION	SULFURIC ACID	STEAM	ACID/BIOMASS TO PRESS	WASH WATER	SOLIDS TO BOILER	BOILER RESIDUE TO LANDFILL	HYDROLYZATE TO STRIPPING	FURFURAL AND ACETIC ACID
WATER	179.09										
HEMI-CELLULOSE	✓ 3.75	9.85	X		1.97			0.63			
CELLULOSE	✓ 37.79	99.21	X		61.51			61.51			
LIGNIN	✓ 10.57	27.76	X		27.76	✓		27.76			
OTHER	✓ 7.50	19.70	X		19.70	✓		19.70	1.97		
ASH	✓ 8.59	22.56	X		22.56	✓		22.56	22.56		
XYLOSE					4.92			0.15		4.78	
GLUCOSE					37.70			1.13		36.57	
FURFURAL					2.95			0.09		2.87	2.87
SULFURIC ACID				10.75		10.75		0.32		10.42	
WATER	✓ 31.79	83.49	✓	321.07	121.97	526.52	137.84	133.03		531.33	
ACETIC ACID										1.34	1.34
CARBON DIOXIDE											
LIME											
GYP SUM											
ETHANOL											
TOTAL	✓ 179.09	100.00	262.56	331.02	✓ 121.97	716.35	✓ 137.84	✓ 266.89	✓ 24.53	✓ 987.31	✓ 4.21

Handwritten notes:

- WATER: 179.09
- HEMI-CELLULOSE: 3.75
- CELLULOSE: 37.79
- LIGNIN: 10.57
- OTHER: 7.50
- ASH: 8.59
- XYLOSE: 4.92
- GLUCOSE: 37.70
- FURFURAL: 2.95
- SULFURIC ACID: 10.75
- WATER: 31.79
- ACETIC ACID: 1.34
- CARBON DIOXIDE: 0.15
- LIME: 0.09
- GYP SUM: 0.32
- ETHANOL: 0.63
- TOTAL: 179.09

Additional handwritten notes:

- WATER: 179.09
- HEMI-CELLULOSE: 3.75
- CELLULOSE: 37.79
- LIGNIN: 10.57
- OTHER: 7.50
- ASH: 8.59
- XYLOSE: 4.92
- GLUCOSE: 37.70
- FURFURAL: 2.95
- SULFURIC ACID: 10.75
- WATER: 31.79
- ACETIC ACID: 1.34
- CARBON DIOXIDE: 0.15
- LIME: 0.09
- GYP SUM: 0.32
- ETHANOL: 0.63
- TOTAL: 179.09

This is same as 8/19.

Table 16. continued.

$31 + 30 = 31$  - sulfuric acid + 0.00144. Use - finding Gypsum  
 $31 + 32 = 33 + 34$  - OK  
 $34 = 35 + 36$  - OK

STREAM NUMBER	28	29	30	31	32	33	34	35	36
STREAM COMPONENT	STEAM	HYDROLYZATE TO NEUTRALIZATION	LIME	HYDROLYZATE GYPSUM SOLUTION	WASH WATER	GYPSUM TO DISPOSAL	HYDROL. TO FERMENTATION	CARBON DIOXIDE	BEER TO DISTILLATION
WDF									
WEMI-CELLULOSE									
CELLULOSE									
LIGNIN									
OTHER									
ASH									
XYLOSE		4.78 ✓		4.78 ✓		0.14 ✓	4.63 ✓	0.46	
GLUCOSE		36.57 ✓		36.57 ✓		1.10 ✓	35.47 ✓	3.55	
FURFURAL			✓						
SULFURIC ACID			10.42 ✓						
WATER		48.42 ✓	331.33 ✓	331.33 ✓	57.16 ✓	45.58 ✓	542.91 ✓		542.91
ACETIC ACID									
CARBON DIOXIDE								17.69	
LIME			7.07 ✓						
GYPSUM				18.29 ✓		18.29			
ETHANOL									18.41
TOTAL	48.42 ✓	583.10 ✓	7.07 ✓	590.97 ✓	57.16 ✓	65.11 ✓	583.01 ✓	17.69 ✓	565.33 ✓

raised this amount  
 ?

= 4835 lb/h  
 where did this disappear

90% converts

OK

$\frac{17.69}{(35.47 + 4.63)} = 49.03$   
 OK

Table 16. continued.

36 + 37 = 38 + 39 ← Food 37 doesn't show up in Products.

$$37 = 40 + 41$$

38 + 44 = 42 + 43 ← Food 44 doesn't show up in Products.

STREAM NUMBER	37	38	39	40	41	42	43	44
STREAM COMPONENT	STEAM	STILLAGE TO EVAPORATOR	ETHANOL TO DEHYDRATION	WATER FROM DEHYDRATION	ETHANOL FROM DEHYDRATION	MAKE-UP WATER TO STORAGE	EVAPORATOR STILLAGE TO BOILER	STEAM
WDF	.....	.....	.....	.....	.....	.....	.....	.....
WMI-CELLULOSE	.....	.....	.....	.....	.....	.....	.....	.....
CELLULOSE	.....	.....	.....	.....	.....	.....	.....	.....
LIGNIN	.....	.....	.....	.....	.....	.....	.....	.....
OTHER	.....	.....	.....	.....	.....	.....	.....	.....
ASH	.....	.....	.....	.....	.....	.....	.....	.....
XYLOSE	.....	0.46 /	.....	.....	.....	.....	0.46 /	.....
GLUCOSE	.....	3.55 /	.....	.....	.....	.....	3.55 /	.....
FURFURAL	.....	.....	.....	.....	.....	.....	.....	.....
SULFURIC ACID	.....	.....	.....	.....	.....	.....	.....	.....
WATER	59.40	541.97 ?	0.94 ✓	0.94	.....	539.29	2.67	180.64
ACETIC ACID	.....	.....	.....	.....	.....	.....	.....	.....
CARBON DIOXIDE	.....	.....	.....	.....	.....	.....	.....	.....
LIME	.....	.....	.....	.....	.....	.....	.....	.....
GYPSUM	.....	.....	.....	.....	.....	.....	.....	.....
ETHANOL	.....	0.55 ✓	17.86 ✓	.....	17.86 ✓	0.55 ✓	.....	.....
TOTAL	59.40	546.53	18.80	0.94	17.86	539.84	6.68	180.64

Basist  
this amount?  
where does  
this appear?

Basist of  
this amount  
where does  
it show up  
150.53/4

37  
95% ethanol - OK



Tennessee Valley Authority Post Office Box 1010, Muscle Shoals, Alabama 35660

September 18, 1991

Mr. Michael C. Mensinger, Manager  
Process Development  
Institute of Gas Technology  
3424 South State Street  
Chicago, Illinois 60616

Dear Mr. Mensinger:

Your letter of September 6, 1991, was received and we appreciate your interest in our work. As discussed in our telephone conversation last week, we will try to provide as much help as needed for you to accurately assess TVA's proposed system for MSW processing and conversion to energy.

We have reviewed your questions and comments on the material balance included in TVA's feasibility study for a potential plant in Southern Indiana. Our comments and corrections are attached. It should be noted that the subject report was prepared by TVA as a pre-feasibility study to determine if a detailed analysis should be conducted to decide the commercial feasibility of applying TVA's technology to a specific Indiana site. The report is general in nature and was prepared prior to the completion of pilot-scale tests necessary to define various aspects of the process.

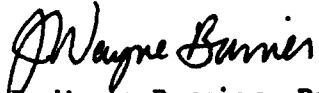
Since the completion of the Indiana report, an in-depth study of the process has begun which will result in a commercial plant design, including detailed material and energy balances, equipment specifications, etc. We are currently working with Kamyr, Inc., Glen Falls, New York to develop this design. Kamyr manufactures the hydrolysis reactor used in the pilot plant and has assisted in development of detailed design premises which will be used in the commercial plant design. These premises (attached) are based on TVA pilot plant and research work. Kamyr visited with TVA's staff in Muscle Shoals to review technology and facilities prior to conducting their analysis of the process. The design being developed by TVA and Kamyr will provide a better basis for your evaluation of the technology than the Indiana pre-feasibility study report. Kamyr has reviewed the comments of the preliminary evaluation report by the Corp of Engineers and has prepared response comments. These are also attached.

2

Mr. Michael C. Mensinger  
September 18, 1991

Please review the attached information and if you have additional questions, please let me know.

Sincerely,

A handwritten signature in cursive script that reads "Wayne Barrier".

W. Wayne Barrier, Program Manager  
Biotechnical Research Department  
National Fertilizer and Environmental  
Research Center

## Response to IGT Questions/Comments on TVA Hydrolysis Report

### Questions 1 and 2:

After preparation of our material balance, new information on moisture contents was received. Changes were made in the material balance from hydrolysis through distillation; so that the amount of water shown in the material balance (85.48) going to hydrolysis is correct. However, adjustments upstream from hydrolysis (i.e., front-end classification) were not made. We made these corrections for the front-end section after the report was issued. The corrected mass balance is attached.

### Question 3:

The steam used in furfural stripping, distillation, and stillage evaporation is for heating purposes. We agree that a condensate stream should also be shown. Steam for furfural stripping is based on literature values for the recovery of furfural from acid hydrolyzates. Steam for stillage processing is based on vendor quotes for the evaporation of water from the stillage stream. Distillation steam was based on commercial ethanol distillation processes.

### Question 4:

The wash water is based on the amount of water necessary to displace the liquid (sugar/acid solution) in the filter cake.

### Question 5:

The concentration of acid in stream 20 is in weight percent. The sulfuric acid concentration of the stream is based on achieving a 2% acid concentration within the reactor, taking into account all sources of water within the reactor (steam, moisture in the feedstock, etc.). On that basis the required acid concentration of the stream is 3.24% not 2.0%. It is agreed that the flow diagram should be updated.

### Question 6:

The acetic acid is formed in the hydrolysis reactor. This quantity was inadvertently left out of the stream leaving the hydrolysis reactor.

### Questions 7 and 8:

Agreed.

## DESIGN CRITERIA

### OVERALL SYSTEM

MSW feed rate to collection/sorting center (As is basis)	T MSW/D	400
MSW % Moisture	%	22

### SORTING AND TRANSFER SYSTEM

Recyclables Removed	T/D	90
Landfill Material Removed	T/D	20
RDF Delivery to Plant	T/D	290

### HYDROLYSIS REACTOR SYSTEM

RDF Feed Rate	T/D	290
Moisture Content	%	28
RDF Density (BD)	lb/ft <sup>3</sup>	10
Reactor Pressure	psig	200
Reactor Temperature	°F	374
Reactor Retention Time (BD RDF)	min	10
Liquor/Solid Ratio at the top of the Reactor		3:1
Sulfuric Acid Charge (on BD RDF)	%	6.0
Sugar Formation (on BD RDF)	lb/T	388
Dissolved Solids Formation (on BD RDF)	lb/T	600

### SULFURIC ACID HANDLING SYSTEM

Sulfuric Acid Concentration (Delivered)	%	93
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### BLOW TANK AND BELT PRESS SYSTEM

Blow Tank Pressure	psig	0
Belt Press Feed Consistency	%	4.0
Flocculant Charge (on BD Sludge)	lb/T	4.0
Belt Press Discharge Moisture Content	%	50
Belt Press Feed Temperature	°F	130
Sugar Loss Across Belt Press	%	15
Excess Limestone Addition (from neutral)	%	100
Sludge Moisture Limit for Incinerators	%	50



### FLOCCULANT HANDLING SYSTEM

Flocculant Concentration in Mix Tank	%	1.0
Flocculant Feed Concentration to Belt Press	%	0.25

### HYDROLYZATE NEUTRALIZATION SYSTEM

Excess Lime Addition (from neutral)	%	10
Target pH after Lime Addition		10.0
Target pH after Sulfuric Addition		7.0
Gypsum Moisture Content	%	35
Drum Washer Dilution Factor	T/BDT	0

### FERMENTATION SYSTEM

Sodium Sulfite Charge to Hydrolyzate (on Sugar)	lb/lb	0.025
Hydrolyzate Temperature to Yeast Preparation Tank, Nutrient Preparation Tank, and Fermentors	°F	92
Fermentor Batch Size	1000 gal	274
Batch Collection Time	hrs	36
Fermentation Time Required	hrs	26
Duration of Peak Fermentation Load	hrs	6
CO <sub>2</sub> Evolution (Average on Sugar)	lb/lb	0.49
Ethanol Yield (Average on Sugar)	lb/lb	0.51
Fermentation pH		7.0

### YEAST HANDLING SYSTEM

Live Yeast Charge (on Sugar)	lb/lb	0.01
Yeast Solution Charge (on Fermentor Contents)	%	2.8

### NUTRIENT HANDLING SYSTEM

Solids Content After Nutrient Batch Preparation	%	20
Potassium Chloride Charge (on Sugar)	lb/lb	0.0075
Urea Charge (on Sugar)	lb/lb	0.05
Phosphoric Acid Charge (on Sugar)	lb/lb	0.01
Killed Yeast Charge (on Sugar)	lb/lb	0.05

### PHOSPHORIC ACID HANDLING SYSTEM

Phosphoric Acid Concentration (Delivered)	%	75
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### CLEAN IN PLACE SYSTEM

1st Rinse Volume	gal	7,000
Sodium Hydroxide Concentration in Cleaning Solution	%	2
Total Circulation of Cleaning Solution	gal	64,000
Circulation Rate of Cleaning Solution	gpm	1,450
2nd Rinse Volume	gal	7,000
Cleaning Solution Temperature	F	175
Total Wash/Rinse Cycle Time	hrs	4

### SODIUM HYDROXIDE HANDLING SYSTEM

Sodium Hydroxide Concentration (Delivered)	%	50
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### DISTILLATION SYSTEM

Ethanol Feed Concentration (Approximate)	%	2.0
Pure Ethanol Feed Rate	gpm	3.7
Cyclohexane Usage (on Pure Ethanol)	%	0.1
Ethanol Product Proof		200

### SOLVENT HANDLING SYSTEM

Gasoline Charge to Ethanol Product (on Ethanol Volume)	%	5
Ethanol Storage Capacity	days	11

MASS BALANCE FOR MSW PROCESSING  
TO FUELS AND CHEMICALS

STREAM NUMBER			1	2	3	4	5	6	7	8
STREAM COMPONENT	QUANTITY OF MSW COMPOSITION (TPD)	MSW COMPOSITION (PERCENT)	MSW TO TROMMEL (TPD)	FERROUS METALS FROM TROMMEL	FINES TO STONER	GLASS FROM STONER	ORGANIC FINES TO HYDROLYSIS	WDF TO PICKING LINES	HDPE TO ROLER	PET TO ROLER
MSW	400.00									
RUBBER & PAPERBOARD		30.98	123.90					123.90		
GLASS		3.90	33.60		39.60	33.66	5.94			
ALUMINUM		1.40	5.60					5.60		
FERROUS METALS/ TIN/STEEL		8.85	35.40	26.55				8.85		
NON FERROUS METALS		0.25	1.00					1.00		
HDPE		2.70	10.80					10.80	10.26	
LDPE		3.20	12.80					12.80		
PET		1.40	5.60					5.60		5.32
OTHER PLASTICS		2.80	11.20					11.20		
RUBBER & LEATHER		2.81	11.22					11.22		
TEXTILES		1.32	7.68					7.68		
FOOD WASTES		3.68	14.70		14.70	0.74	13.97			
WOOD		3.60	14.40		14.40	0.72	13.68			
YARD WASTES		2.55	10.20		10.20	0.51	9.69			
MISC. INORGANICS		2.10	8.40		8.40	0.42	7.98			
WATER		21.88	87.50		28.32	0.91	28.00	58.58		
TOTAL	400.00	100.00	400.00	26.55	116.22	36.96	79.26	257.23	10.26	5.32

MASS BALANCE CONTINUED

STREAM NUMBER	9	10	11	12	13	14	15	16	17	18	19
STREAM COMPONENT	LDPE TO LANDFILL	NON-FERROUS TO STORAGE	PAPER TO SHREDDER	PICKING LINES TO ALUMINUM SEPARATOR	ALUMINUM TO RYLER	MDF TO MAGNET	METALS FROM MAGNET	MAGNET RESIDUE TO AIR CLASSIFIER	HEAVIES TO LANDFILL	LIGHTS TO SHREDDER	MDF TO HYDROLYSIS
MSW	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
PAPER & PAPERBOARD	-----	-----	74.34	43.56	-----	43.56	-----	43.56	2.48	47.08	121.42
GLASS	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	5.94
ALUMINUM	-----	-----	-----	5.60	5.04	0.56	-----	0.56	-----	0.56	0.56
FERROUS METALS/ TIN/STEEL	-----	-----	-----	8.85	-----	8.85	7.08	1.77	1.42	0.35	0.35
NON-FERROUS METALS	-----	0.35	-----	0.05	-----	0.05	0.05	-----	-----	-----	-----
HDPE (milk)	-----	-----	-----	0.54	-----	0.54	-----	0.54	-----	0.54	0.54
LDPE	12.80	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
PET (soda)	-----	-----	-----	0.28	-----	0.28	-----	0.28	-----	0.28	0.28
OTHER PLASTICS	-----	-----	-----	11.20	-----	11.20	-----	11.20	10.64	0.56	0.56
RUBBER & LEATHER	-----	-----	-----	11.22	-----	11.22	-----	11.22	9.35	1.87	1.87
TEXTILES	-----	-----	-----	7.68	-----	7.68	-----	7.68	5.44	2.24	2.24
FOOD WASTES	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	13.97
WOOD	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	13.68
YARD WASTES	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	9.69
MISC. INORGANICS	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	7.98
WATER	-----	-----	28.34	30.24	-----	30.24	-----	30.24	3.11	27.14	83.48
TOTAL	12.80	0.35	102.68	125.22	5.04	120.18	7.13	113.05	32.43	80.63	262.57

## MASS BALANCE CONTINUED

STREAM NUMBER				20	21	22	23	24	25	26	27
STREAM COMPONENT	QUANTITY OF DRY WDF (TPD)	PERCENTS OF WDF (WET)	BIOMASS FEEDSTOCK	SULFURIC ACID SOLUTION	STEAM	ACID/ BIOMASS TO PRESS	WASH WATER	SOLIDS TO BOILER	BOILER RESIDUE TO LANDFILL	HYDROLYZATE TO STRIPPING	FURFURAL AND ACETIC ACID
WDF	179.09	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
HEMI- CELLULOSE	-----	3.75	9.85	-----	-----	0.63	-----	0.63	-----	-----	-----
CELLULOSE	-----	37.79	99.21	-----	-----	61.51	-----	61.51	-----	-----	-----
LIGNIN	-----	10.57	27.76	-----	-----	27.76	-----	27.76	-----	-----	-----
OTHER	-----	7.50	19.70	-----	-----	19.70	-----	19.70	1.97	-----	-----
ASH	-----	8.59	22.57	-----	-----	22.57	-----	22.57	22.57	-----	-----
XYLOSE	-----	-----	-----	-----	-----	4.92	-----	0.15	-----	4.78	-----
GLUCOSE	-----	-----	-----	-----	-----	37.70	-----	1.13	-----	36.57	-----
FURFURAL	-----	-----	-----	-----	-----	2.95	-----	0.09	-----	2.87	2.87
SULFURIC ACID	-----	-----	-----	10.75	-----	10.75	-----	0.32	-----	10.42	-----
WATER	-----	31.79	83.48	321.07	121.97	526.52	136.30	133.03	-----	529.79	-----
ACETIC ACID	-----	-----	-----	-----	-----	1.34	-----	-----	-----	1.34	1.34
CARBON DIOXIDE	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
LIME	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
GYPSON	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
ETHANOL	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
TOTAL	179.09	100.00	262.56	331.82	121.97	716.35	136.30	266.89	24.54	585.76	4.21

# MASS BALANCE CONTINUED

STREAM NUMBER	28	29	30	31	32	33	34	35	36
STREAM COMPONENT	STEAM	HYDROLYZATE TO NEUTRAL- IZATION	LIME	HYDROLYZATE GYPSUM SOLUTION	WASH WATER	GYPSUM TO DISPOSAL	HYDROL. TO FER- MENTATION	CARBON DIOXIDE	BEER TO DISTILLATION
WLF									
HEMI- CELLULOSE									
CELLULOSE									
LIGNIN									
OTHER									
ASH									
XYLOSE		4.78		4.78		0.14	4.63		0.46
GLUCOSE		36.57		36.57		1.10	35.47		3.55
FURFURAL									
SULFURIC ACID		10.42							
WATER	48.42	523.79		523.79	57.16	45.58	541.36		541.36
ACETIC ACID									
CARBON DIOXIDE								17.69	
LIME			7.87						
GYPSUM				18.29		18.29			
ETHANOL									18.41
TOTAL	48.42	581.56	7.87	583.43	57.16	65.11	581.47	17.69	563.78

MASS BALANCE CONTINUED

STREAM NUMBER	37	38	39	40	41	42	43	44
STREAM COMPONENT	STEAM	STILLAGE TO EVAPORATOR	ETHANOL TO DEHYDRATION	WATER FROM DEHYDR.	ETHANOL FROM DEHYDRATION	MAKE-UP WATER TO STORAGE	EVAPORATOR STILLAGE TO BOILER	STEAM
WDF								
HEMI- CELLULOSE								
CELLULOSE								
LIGNIN								
OTHER								
ASH								
XYLOSE		0.46					0.46	
GLUCOSE		3.55					3.55	
FURFURAL								
SULFURIC ACID								
WATER	53.40	540.42	0.94	0.94		537.75	2.67	180.64
ACETIC ACID								
CARBON DIOXIDE								
LIME								
GYPSSUM								
ETHANOL		0.55	17.86		17.86	0.55		
TOTAL	53.40	544.93	18.80	0.94	17.86	538.30	6.68	180.64

September 17, 1991

Mr. J. Wayne Barrier  
Program Manager  
Biotechnical Research Department  
Tennessee Valley Authority  
P.O. Box 1010  
Muscle Shoals, AL 35660

Dear Wayne

You have asked for Kamyr's comments on the report which you received from the Corps of Engineers, "ANALYSIS OF THE TVA/TITAN DILUTE ACID AND UAH/REDSTONE ARSENAL ENZYMATIC ETHANOL PRODUCTION FACILITY PROPOSALS", which was sent to you on August 27. However, it appears that it may have been written, or at least researched, somewhat earlier, as things have been moving fast and some of the information included on the TVA process is now out of date. It also looks as though some of the questions raised are based on conflicting information, some of which came from reports several years old. Our comments will be based on the most recent information we have received from you and from other sources whom we have contacted in carrying out the preliminary engineering for a 400 T/D MSW-to-Ethanol plant based on your dilute acid hydrolysis process.

The report is slightly confusing in that it addresses both the TVA and UAH processes together, and in some cases it is a little difficult to tell which process is being referred to. The discussion below only refers to the report as it applies to the TVA process, as we have no experience with the UAH process, and cannot judge its merits at this time.

There are quite a number of points raised in the report, covering economics, technology, accuracy of base data, calculations, and political questions. Rather than cover each and every line of the report, we will specifically address each of the conclusions and recommendations, as these incorporate and are based on the more extensive information contained in the body of the report. We believe some of these conclusions are correct, but disagree with some of them, and will try to spell out exactly where we agree or disagree. If you would like for us to review anything in more detail in any part of the report, please let me know.

Conclusion one states:

"The MSW to ethanol processes developed by TVA and UAH are technically feasible. Bench scale testing and moderate pilot batch testing have provided sufficient information to design a continuous pilot plant. Neither process is developed enough to design a commercial scale demonstration plant."



As noted above, we cannot address any aspects of the UAH process since we have not analyzed any of the ethanol from waste processes except for yours. However, it is probably worth noting that TVA has looked at all of the known processes in considerable detail, and you have concluded that the dilute acid hydrolysis process is the best one available today. It addresses both the landfill problem and alternative fuels production, which many other processes do not. You are also convinced of both the technical and economic feasibility of this process, as are we.

Addressing conclusion one, and the others, only in light of their reference to the TVA process, we definitely agree that it is technically feasible. However, we do not agree with the final sentence. Based on the information we have received from you, and the other testing planned for the near future, there is certainly adequate information available to design a commercial scale demonstration plant. We have carried out such a design for a 400 T/D plant, and have sent you our estimate for the cost of this plant together with the flowsheets and layout drawings on which the cost estimate is based. Further, as stated in my letter to you of September 10, we are prepared to commit considerable money and resources of our own to help see that the plant is indeed built and successfully placed in full operation in the near future. I think that this is the most convincing proof that we can provide that we believe in the commercial potential of the TVA process.

Conclusion two states:

"Significant questions exist about the economic viability of the MSW to ethanol processes. A re-evaluation of the process economics should be undertaken including a cost comparison with competing resource recovery and MSW disposal technologies."

Here again, we cannot answer for processes other than yours. We have given you our estimated capital costs for the plant, which we believe to be correct. I agree that some of the earlier capital estimates were overly optimistic. We believe our estimates to be very realistic, and we have considerable experience in estimating and building large, complex, turnkey plants at firm prices, and we are still in business after having done this for many years.

With respect to the relative economics of your process versus others, we defer to TVA's judgment and expertise. The same applies to the MSW disposal techniques, on which you have done considerable research.

Conclusion three states:

"Both processes initially were aimed to convert woody and agricultural biomass to liquid fuels during the period of high oil prices, and were later moved into conversion of MSW when environmental issues became the center of concern."

This is certainly true, and the earlier research gave a very good foundation for the present MSW based work. The largest differences are economic rather than technical, and the economics are certainly much more favorable when someone is paying you to take the raw material, rather than your paying them.

**Conclusion four states:**

"Both processes adapted equipment and technologies from a variety of industrial practices ranging from paper production to meat rendering. The UAH process has been fully tested in batch mode from MSW to ethanol. The TVA process has been fully tested from MSW through dilute acid hydrolysis. Both lack the integrated engineering approach needed for a continuous process, with significant questions remaining about the types of equipment used, their operational layout and the O&M procedures required for a continuous process."

It is very true that the TVA process utilizes equipment and technology used extensively in other industries, including the paper industry. In fact, this is one aspect of the process that greatly reduces its risk. Almost every single one of the unit operations, and every piece of the equipment, has been used extensively in other applications. The only real difference is in the use of the zirconium metallurgy in the hydrolysis section of the plant. Although the hydrolysis system is the heart of your process, it is actually only a small part of the overall plant, less than 25%. The hydrolyzer feed equipment, the reactor, and the discharge equipment are all very conventional in design, except for the metallurgy. You have conducted extensive pilot plant tests on a variety of feed stocks over a wide range of operating conditions, far wider than would be encountered in a commercial operation, and the zirconium has worked very well.

In terms of the extent of the testing, the pilot plant operations have now been extended through the fermentation stage. Our understanding is that the fermentation trials, which started in April, have been fully successful in every respect, and have met or exceeded all expectations.

As noted elsewhere in the report, the ethanol distillation and drying are quite conventional. We obtained quotes for package systems for these, and they would be brought in as preassembled units and set on our foundations. The distillation unit vendor chosen has installed a number of such systems using identical technology.

With regard to the "integrated engineering approach", this is much further along than it was several months ago. This is part of the package that Kamyr brings to the table. We have been designing and building large turnkey fiberline plants in the pulp and paper industry, using our own proprietary technology, for a number of years. These plants are similar in many ways to the proposed TVA MSW-to-Ethanol plant. We are certainly far enough along with the overall plant balances, flowsheets, layouts and estimates to say that we can build such a plant, and that it will operate successfully.

I should point out that we don't make judgments like this randomly or without due consideration. Kamyr is known worldwide for innovation and for our ability to bring new technology to market. We have decades of experience in evaluating the risks in commercializing new processes and equipment, and you can check with any of our customers in this regard and get full confirmation on this point. They will also confirm that the Kamyr name is synonymous with continuous processing.

**Conclusion five states:**

"Economic analysis of both processes appear to rely on some optimistic estimates for costs (both capital and O&M), energy balance and mass balance. Some assumptions made appeared to be unrealistic such as "steam is free"

(instead of \$3-\$4/ton MSW) for the UAH process, both processes assume a high energy credit for residue burning (8600 BTU/lb, gross instead of 3800 BTU/lb, net after the 50% water has been driven off). The UAH process assumes you can convert scotch marine fire tube boiler to burning solid fuel. This conversion would probably not be cost effective based on coal conversion experiences. TVA assumes you can build a solid fuel burning circulating fluidized bed boiler for \$4.53/MBTU/hr capacity (really costs \$170/MBTU/hr capacity), and TVA blames low ethanol yield (12-16 gallons/ton MSW) on waste paper recycling by the Humbolt TN RDF plant. It should be noted that waste paper recycling will be more prevalent in the future, so this will be more typical than the 30 gal/ton figure used."

We agree that some of the earlier capital cost estimates were low. We feel that ours are well grounded and realistic. We also feel that our energy and mass balances are correct, with the exceptions that: (a) better residue incineration data is needed and (b) residue drying requirements need to be more fully defined. This work is now in progress, and your residue burn tests are starting at Hazen Research this week. Pyropower, the incineration division of our parent company, Ahlstrom, would like to get this data as soon as it is available in order to confirm their preliminary calculations. If additional residue drying is required prior to incineration there are several commercial systems we can easily incorporate into the existing layouts and flowsheets; but the changes must be taken into account in the balances and in the other affected areas. This should not present a problem.

With regard to the ethanol yields, we have chosen conditions which correspond to 13.3 gallons of ethanol per ton of raw MSW, based on a number of conversations with Jacqueline Broder. She in turn has looked at data from a number of areas of the country, and we believe this figure to be realistic and conservative. I'm not sure where the 30 gal/ton figure comes from, but it looks more like it is based on dry RDF than on raw MSW, so maybe there was some mixup involved.

I would like to address the waste paper recycling question. We are involved in building both virgin fiber pulp mills and recycle fiber plants, having entered the recycle market only recently due to the rising demand for these plants. Fibers can only be effectively recycled 3 to 4 times. In this country we are now at about a 25% recycle rate. Even if we can achieve the targeted rate of 40% by the year 2000, there will still be a net increase of fiber to landfill because of the projected growth in total fiber demand. And, the target of 40% recycle would require construction of a new 300 T/D recycle fiber plant every 18 days from now until the end of the century, which is not a very realistic proposition. The bottom line is that the total amount of cellulosic material going to landfill or plants such as the proposed TVA MSW-to-Ethanol plant is not going to decrease, even with the rising rate of recycle fiber usage.

Conclusion six states:

"The TVA process time, temperature, and mixture conditions were different from the bench scale tests (which specified 160°C, 15 minute retention time and 5:1 liquid to solids ratio) to the pilot plant tests (which specified 190°C, 4-6 minute retention time and 3:1 liquid to solids ratio). The UAH process operating conditions appear to be fitted to the available equipment and energy sources, instead of analyzed to reflect possible scale up to commercial systems."

A wide variety of conditions were checked in both the lab and pilot plant. The design conditions we have used, 190°C for 10 minutes, reflect all of the work you have

done to date and are your choice for the best design point. However, should you choose to change these conditions, the effect on the process, the capital cost, and the operating cost would be essentially negligible based on the range of conditions which could be reasonably expected.

Conclusion seven states:

"DOE does not consider the TVA dilute acid process ready for a scale up to 100 tpd for hardwood to ethanol pilot plant. DOE considers any plant for converting waste paper/RDF to ethanol to be even more risky because of the metals found in MSW which are toxic to the yeast fermentation process. Considerable amount of additional research needs to be done before building the 100 tpd TITAN/TVA dilute acid hydrolysis plant."

Conclusions based on earlier hardwood work may not really be relevant here, but in any event, we are prepared to design and build a commercial size MSW-to-Ethanol plant whenever you say.

The potential toxic effect of metals on the yeast fermentation process is a valid point, and I understand that TVA also had some serious concerns about this when you started the MSW work. However, you have seen absolutely no problem with this, even when processing raw MSW. It appears that the neutralization step prior to fermentation precipitates the toxic metals and locks them into the residue/solids side of the process, and analysis of the various process streams in both lab and pilot plant work confirms this hypothesis.

With regard to the final sentence of conclusion seven, we agree that some additional engineering design data is needed. However, with due diligence the remaining open questions can be answered in a few months using existing facilities. There will still be some unknowns even if you build and operate a 20 ton/day pilot plant, as later recommended in this report. In fact, there will still be some unknowns when you start to build plant number 10 or plant number 100. At least there are when we build new pulp mills, and the pulp and paper industry has been around for a long, long time.

In looking at the risks of proceeding with a full scale demonstration plant, I would say they are almost entirely quantitative rather than qualitative. By this I mean that without question such a plant will work, but it might well initially produce at less than the full design rate; typically because of undersizing some piece of equipment such as a filter or a pump, for example. This can be avoided by a lot of over design, but excessive over design is not a very cost effective way to reduce risk, and excessive safety factors tend to be perpetuated in future designs for a long time. A much better alternative is to leave room in the layout for expansion or addition of more modules of any equipment which may prove to be too small or otherwise inadequate.

One concern that should be addressed, as with any new plant that must be able to handle a continuing day-to-day influx of feed material with high reliability, is what to do if the plant is shut down for repairs or modifications. These will be necessary even if the plant is built after 10 more years of pilot plant testing. What we suggest is to build the demonstration plant as proposed in our letter to you of August 23, with the incinerators designed and sized to handle the RDF directly after sorting, and with the Ethanol plant installed as a side-stream, or alternate processing route. The ability to utilize this arrangement for the first large scale system at a comparatively low cost would be one of the most significant benefits of building a demonstration plant, and the

effects of any possible downtime resulting from the Ethanol plant could be effectively avoided. In practice, this can be accomplished for very little extra cost, since in a typical RDF incineration plant the incinerators are modular, with one unit serving as a spare. This spare capacity can also be utilized to pick up the extra load in the event of an ethanol plant outage.

Conclusion eight states:

"The 100 tpd TITAN/TVA plant will not be economical. Preliminary cost estimates by TVA show the minimum economical plant size to be 500 tpd."

We agree. In fact we would put the number at 1000-2000 tons/day for a normal installation. Some special cases in areas where high tipping fees are required might justify plants in the 500 ton/day range.

However, in view of conclusion nine, listed next, it should be pointed out that these figures are in tons/day of MSW, not paper. There may have been some confusion here.

Conclusion nine states:

"To generate 500 tpd of waste paper, the plant would have to take in over 1600 tpd of MSW (over 800,000 population base). To generate 100 tpd of waste paper, the plant would have to take in over 300 tpd of MSW (over 160,000 population base). Most Army installations generate less than 50 tpd of MSW. The 27 tpd UAH/Redstone Arsenal plant would have to take in over 250 tpd of MSW (over 130,000 population base)."

I believe TVA's figures are based on EPA estimates of 5 pounds of MSW per person per day in urban areas. On this basis a 1000-2000 ton/day MSW plant would require a population base of 400,000 to 800,000 people. With regard to the total amount of MSW generated by various Army bases we must obviously defer to the expertise of the report authors.

However, it has been our understanding from the beginning that there was no intention to put a commercial size plant on an Army base and use it only for Army generated MSW since this would obviously not be economical. Rather, the proposed Army base location was chosen on the basis of also handling MSW from nearby communities in the interest of promoting public relations and community approval and acceptance for the Army.

Conclusion ten states:

"Both the TITAN/TVA and UAH/Redstone processes are still under development and are not ready for commercial application. The pilot plant process for either option will require a lot of intervention by highly skilled labor and researchers, neither of which are typically found on Army installations."

First, as noted above, we at Kamyr are completely convinced that the TVA process is ready for scaleup to a plant in the four hundred ton per day range. The technological challenge of designing a plant for one to two thousand tons per day is very little different from that of designing one for 400 tons/day, and we have already been through the preliminary design for a plant of this size. You have received the estimates, flowsheets, layouts, etc. that are a part of this engineering design package.



You also have our letter of September 10 in which we state that Kamyr is willing to commit considerable money and resources of our own, based on our firm belief in exactly this; that the TVA process is at the point where a full size demonstration plant is technologically viable, with full scale commercialization to follow in short order.

Second, regarding the initial requirements for researchers and skilled labor, this is a valid point. Both TVA and Kamyr will need to devote a number of qualified people to the demonstration plant project for some years in order to ensure its success. We would expect to do this on any project involving new technology, as this is the only way to make new technology work. Also, a considerable amount of work will have to be carried out at the demonstration plant site in order to obtain detailed design data for future, larger scale plants. This will also require extensive TVA and Kamyr involvement. This said, it should also be noted that Kamyr has installed continuous digesters and bleach plants all over the world, in many third world countries, and in places where the operators cannot read or write. Once fully trained, they can still be excellent operators, who take great pride in doing a good job. The MSW-to-Ethanol plant, once past the initial stages, will be a relatively easy plant to operate and maintain. For any difficult areas, welding of zirconium for example, contract maintenance is readily available anywhere in North America.

Moving on to the recommendations, recommendation one states:

"Prior to undertaking any pilot plant programs, the entire TVA and UAH process should be re-evaluated for economics, material balance and energy balance. At a minimum, three scenarios should be evaluated: best case, worst case and expected case. To determine the relative value of the TVA and UAH processes, they should also be compared with competing technologies such as recycling, incineration, MSW gasification, composting, conventional landfill, etc. If after performing these studies the TVA and UAH processes still seem to be economically viable, investment in a continuously operating pilot plant should be considered."

We have already done a number of the recommended calculations for the "expected case" Scenario. We are now computerizing the heat and material balances, which should be finished by October. I suggest that after this is completed we feed in some best and worst case assumptions and forward the results to you, and you can have your economists review the economics. This is certainly something that needs doing at some point along the way.

The comparison with other processes is also much more in TVA's area of expertise than ours, based on your past research in this area.

As noted earlier, we at Kamyr feel that construction and prolonged operation of a 20 ton/day pilot plant is not required, and that we should proceed directly to the 400 ton/day scale. However, the existing pilot plant facilities should be kept operational to provide fuel for burn tests, explore various operating parameters, carry out trials on other feed stocks, and test out evolutions in the hydrolysis process on a pilot plant scale.

Recommendation two states:

"At their present state of development, both the TVA and UAH processes are not ready for scale up to economically viable commercial applications. TVA has the needed personnel and equipment to conduct an expanded pilot plant

study. Both processes are advanced enough to serve as a basis for detailed engineering and technology development studies that could lead to the design, construction and operation of a continuous operation pilot plant. It is recommended that the current TVA pilot batch plants for both acid hydrolysis and fermentation be expanded to a 20 tons/day continuous pilot plant that takes unprocessed MSW in and puts out commercial grade ethanol (170 or 200 proof). It is also recommended that the 50% moisture residue be burned in a TVA or other solid fuel plant (fluidized bed, pulverized coal and stoker technology should be tested). The pilot plant system should be adaptable enough to incorporate the most advantageous aspects of the UAH process and other processes for waste classification, cellulose hydrolysis, solids dewatering, and alcohol fermentation. A 2 year intensive pilot plant study should provide sufficient information to adequately design and evaluate a commercial scale demonstration plant."

Our feelings on the viability of commercial operation, and the need for the 20 ton/day pilot plant, are noted above. We emphatically agree with the report on the need for the burn tests.

With regard to a 20 T/D MSW handling and classifying pilot facility, this is probably better set up some place in parallel with an existing plant. Feed stock and product disposal problems are then minimized, and equipment capital requirements are also reduced. We feel that this project, although probably justifiable on its own merits, is not necessary in order to proceed with the MSW-to-Ethanol demonstration plant, and should not be allowed to affect the demonstration plant schedule. MSW collection and sorting techniques have already been proven in numerous installations. Certainly more and more improvements will continue to be made, but this is true in anything from computer systems to automobile manufacturing, and next year's model can always be a little bit further along. The MSW-to-Ethanol plant is not dependent on further evolution in MSW handling technology. Any further improvements in this area can almost always be retrofitted into this type of plant where worthwhile, and can be incorporated from the start in future commercial plants.

Recommendation three states:

"Both processes are not developed enough to be applied to an Army installation, and both processes are targeted at cities with a population of over 130,000. The only future for these technologies in solving Army MSW problems is as part of a regional waste management plan. The army should not participate until these technologies are more mature, and such a regional opportunity exists."

As noted earlier, we feel that the technologies are already more than far enough along to build a full scale plant. However, we agree that to achieve economic operation, any Army base would have to combine the Base's MSW with that from nearby urban areas to produce economically viable amounts of plant feedstock. It was our understanding that this was the approach being proposed, and we certainly hope to see it succeed.

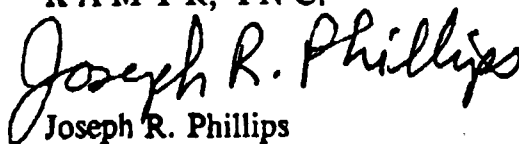
**KAMYR**

In summary, in addition to addressing the conclusions and recommendations of the report, there are some items in the body of the report that warrant discussion. Again, we agree with some of these, and disagree with some. However, it would probably be better to address these in some sort of technology and design review meeting with TVA and the appropriate people from the Corps of Engineers. Such a meeting might also clear up any points regarding outdated or conflicting information. If you feel that this would be appropriate, perhaps you could set up such a technical meeting between say TVA, the Corps of Engineers, Titan, and ourselves. We would be more than happy to attend and to do whatever we can to help move the project along. If you decide to proceed in this way, please let me know when and where the meeting will be, and what you would like for us to be prepared to discuss, and by all means give me a call if we can be of any further assistance.

Best regards.

Yours truly,

K A M Y R, I N C.



Joseph R. Phillips  
Vice President  
Research & Development

JRP:ehp

1APLTR57





Headquarters

September 6, 1991

Professor Michael H. Eley, Ph.D.  
Biological Sciences & Biotechnology  
Kenneth E. Johnson Research Center  
The University of Alabama in Huntsville  
Huntsville, Alabama 35899

Dear Professor Eley:

As you are aware from your conversations with Dr. Anil Goyal, the U.S. Army Construction Engineering Research Laboratory (CERL) has contracted the Institute of Gas Technology (IGT, Chicago) to conduct a process and economic evaluation of the Ethanol Production Facility feasibility study, prepared by Leo A. Daly and based on the enzymatic hydrolysis process.

IGT has made a preliminary review of the material included in the above-referenced feasibility study and has prepared a list of questions and comments that arose during the evaluation. Most of the questions or comments relate to the material balance or to processing steps or conditions that have not been explicitly identified. The list of questions and comments is attached.

It would greatly facilitate IGT's completion of the process review if you, or your designee, could answer the questions and provide the needed information. As you can imagine, time is of the essence.

If the information could be faxed to IGT (facsimile no. 312-567-5209) by September 18 or earlier, it would be greatly appreciated. Additional questions may be faxed to you as they arise.

Thank you in advance for your cooperation and prompt attention to this request. Please call me at (312)-567-3730, or Dr. Goyal at (312)-567-5759 if you have questions about any of the attached comments.

Sincerely,

Michael C. Mensinger  
Manager  
Process Development

cc: Dr. Mike Lin, CERL  
A. Goyal  
C. Blazek

Institute of Gas Technology

Headquarters

3424 South State Street  
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C-31

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**Questions/Comments concerning the Leo A. Daly Ethanol Production Facility feasibility study.**

Please refer to the enclosed material balance and process flow diagram for specific questions/comments.

1. Each process stream should be given a unique stream number for proper identification. Streams 3, 4, 5, 6, 20, 25, 37, and 45 have been used more than once.
2. Some important streams, including those leaving the system, have not been numbered (such as fusel oil recovered from the distillation step, steam entering the fuel dryer, and condensate leaving the fuel dryer).
3. Some material balances are violated. A copy of the flow sheet showing the corrected values is attached. Corrected material balance tables are also included.
4. The fusel oil make for the 16-ton/day case should be 557 pounds instead of 443 pounds. Thus the fusel oil produced is 557 and 573 pounds for the 16 and 27 ton/day cases, respectively. These values are not in the expected proportion of 27/16 or 1.6875.
5. The "net daily material balance" table has errors. The steam and water feeds shown are incorrect. All recycle streams are internal streams and do not leave the system. Recycle streams should not be included as "OUT" streams. There are two condensate streams leaving the system and these must be included. Fusel oil make is not included.
6. In the "Solid Fuel Recovery" balance table, the steam entering the subsystem and condensate leaving the subsystem are not shown.
7. In the "Ethanol Recovery" balance table, the amount of "Still Bottom" leaving the subsystem is shown incorrectly.
8. The enzyme feed rates to the upper Slurry Tank are different (176339 and 190958 pounds), although the hammer-milled paper feed rate is the same for the two cases. These should be identical. The enzyme feed to the lower slurry tank should be in the ratio of 1:2; the same as that of the feeds to the steam processor.
9. The system requires 46895 and 67795 pound/hour of steam for the two cases, respectively. How is the steam generated?
10. In the Fuel Dryer, 1 pound of steam is added to evaporate 1 pound of water. There should be some allowance for heat losses and inefficiency.
11. The compositions of the two types of feed are not specified.
12. The temperature of the various streams are not specified. Accurate energy balances cannot be made without this information.

Questions/Comments concerning the Leo A. Daly Ethanol Production Facility feasibility study.

Please refer to the enclosed material balance and process flow diagram for specific questions/comments.

13. The basis of designs for various stages are not provided, such as residence times, concentrations, temperatures, pressures, etc. Items related to these factors cannot be checked.
14. What is the basis for the cost estimate of the Pervaporation Subsystem? Why was pervaporization selected over more conventional dehydration processes?
15. What are the bases for the hydrolysis reactors and the fermentation vessels?

### **The University of Alabama in Huntsville Process for Conversion of Cellulose to Ethanol**

Waste paper is received and processed to pulp, the paper products and to separate the non-cellulosic materials from the paper pulp. The cellulosic pulp is added to an enzyme solution. After 24 hours in enzyme hydrolysis, the mixture is passed through a solids recovery system to remove the unreacted suspended solids. These solids are washed to remove residual sugar and enzymes, and the solids are dried and processed into solid fuel pellets. The combined liquids are processed by ultrafiltration to separate and concentrate the cellulase enzymes for recycle in the process. The resultant dilute sugar solution, is then processed by reverse osmosis to concentrate the sugar and produce ultra-pure water which is recycled in the process. The concentrated sugar solution is fermented by the addition of yeasts. The carbon dioxide gas produced during fermentation is collected and transported to a carbon dioxide liquification unit. The liquified carbon dioxide is sold as a by-product. The "beer" that is produced by fermentation is then processed to recover the yeasts of which a small portion is recycled and the remainder is sold as a by-product. After yeast recovery, the "beer" is processed in a combination distillation and pervaporation system to remove the water; to separate a small quantity of fusel oil, also sold as a by-product, and to yield the anhydrous ethanol product.

The University of Alabama in Huntsville process for conversion of cellulose to ethanol contains some proprietary information that will not be disclosed in this forum. Sufficient information is provided in this brief narrative and the accompanying flow schematics and material balances that the overall process can be followed and understood without disclosure of any proprietary details. Also accompanying these documents are an equipment layout for a renovated Building 5410 located at Redstone Arsenal, AL, a capital equipment list, and estimated capital equipment costs for a 16 and a 27 ton per day clean paper to ethanol production facility.

[illegible]

(3)

**NET DAILY MATERIAL BALANCE**  
(Based on 16 & 27 Ton/Day)

IN		16 TON	27 TON	
①	Waste Paper	22,000 ✓	44,000 ✓	lbs/day
⑪	Hammermilled Paper	10,000 ✓	10,000 ✓	lbs/day
⑤ + ⑦	Steam $20900 + 25995$	46895 ✓	67795 ✓	lbs/day
②① + ②②	Water $41800 + 25995$ $26600 + 1577$	41,935 ✓	78,285 ✓	lbs/day
⑬	Enzyme $44547 + 2669$	28,177 ✓	47,216 ✓	lbs/day
②⑤	Yeast	152 ✓	257 ✓	lbs/day
②⑥	Ammonia	43 ✓	75 ✓	lbs/day
②⑦		871 ✓	1,508 ✓	lbs/day
		103,179 ✓	181,341 ✓	lbs/day
		108138 ✓	170851	

OUT		16 TON	27 TON	
②③ + ②④	Water Recycled	14,143 ✓	27,743 ✓	lbs/day
②③ + ②④	Steam Condensate Recycled	28,350 ✓	51,115 ✓	lbs/day
②⑤ + ②⑥	Evaporated Moisture	26,324 ✓	44,422 ✓	lbs/day
②⑦	Solid Fuel Product	20,728 ✓	35,051 ✓	lbs/day
②⑧	CO <sub>2</sub> Product	6,334 ✓	10,687 ✓	lbs/day
②⑨	Yeast Product (70% Moisture)	1,054 ✓	1,784 ✓	lbs/day
③①	Fusel Oil Product	57 ✓	673 ✓	lbs/day
③②	200-Proof Ethanol Product	6,246 ✓	10,539 ✓	lbs/day
		103,179 ✓	181,341 ✓	lbs/day
		108029	170851	
		1140	0	

IN - OUT

(4)

STEAM CLASSIFICATION

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
1 Paper	- 22,000 ✓	44,000 ✓	lbs/day
2 Hotwater	- 19,800 ✓	39,600 ✓	lbs/day
3 Steam	- <u>20,900</u> ✓	<u>41,800</u> ✓	lbs/day
	62,700 ✓	125,400 ✓	lbs/day (4)

<u>OUT</u>	<u>16 TON</u>	<u>27 TON</u>	
4 Processor Steam Vent	- 7,315 ✓	14,630 ✓	lbs/day
8 Cellulosics to Conversion	- <u>55,385</u> ✓	<u>110,770</u> ✓	lbs/day
	62,700 ✓	125,400 ✓	lbs/day (5k)

2 HOT WATER

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
4 Steam Vent Condensate	- 1,759 ✓	3,980 ✓	lbs/day
5 Still Bottoms	- <u>18,041</u> ✓	<u>35,620</u> ✓	lbs/day
	19,800 ✓	39,600 ✓	lbs/day
<u>OUT</u>			
2 Hot water to steam processor	19800	39600	lbs/day

5

# SOLID FUEL RECOVERY

IN	16 TON	27 TON	
16 Hydrolysis Slurry	- 1,219,817 ✓	2,064,661 ✓	lbs/day
21 Solids Wash Water	- 89,150 ✓	150,523 ✓	lbs/day
Act → Steam →	25,995	?	
↑ This steam should be added	1,308,967 ✓	2,215,184 ✓	lbs/day (ok)
OUT	16 TON	27 TON	
25 Fuel Moisture Evaporated	- 25,995 ✓	43,867 ✓	lbs/day
26 Solid Fuel Product	- 20,728 ✓	35,051 ✓	lbs/day
29 Liquids to Ultrafilter	- 1,262,244 ✓	2,136,266 ✓	lbs/day
Condensate →	25,995	?	
	1,308,967 ✓	2,215,184 ✓	lbs/day (ok)

# 21 SOLIDS WASH WATER

IN	16 TON	27 TON	
4 Processor Steam Vent	- 5,556 ✓	10,650 ✓	lbs/day
5 Still Bottoms	- 56,994 ✓	95,326 ✓	lbs/day
20 Fresh Water Make-up	- 26,600 ✓	44,547 ✓	lbs/day
OUT	89,150 ✓	150,523 ✓	lbs/day
21 water to solids wash	89,150	150,523	lbs/day



6

# CELLULOSICS TO HYDROLYSIS

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
8 Processed Cellulosics	- 55,385 ✓	110,770 ✓	lbs/day
11 Hammermilled Paper	- 10,000 ✓	10,000 ✓	lbs/day
12 Dilute Enzyme Solution	- <u>1,154,432</u> ✓	<u>1,943,891</u> ✓	lbs/day
	1,219,817 ✓	2,064,661 ✓	lbs/day (a)

<u>OUT</u>	<u>16 TON</u>	<u>27 TON</u>	
16 Slurry to Hydrolysis	- 1,219,817 ✓	2,064,661 ✓	lbs/day (a)

## 12 DILUTE ENZYME SOLUTION

*Enzyme Recycle Storage*

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
<i>with filter</i> 6 R. O. Water <i>Reverse Osmosis</i>	- 1,089,521 ✓	1,833,662 ✓	lbs/day
14 U. F. Concentrate	- 64,759 ✓	109,972 ✓	lbs/day
13 <del>15</del> Dry Make-up Enzyme	- <u>152</u> ✓	<u>257</u> ✓	lbs/day
	1,154,432 ✓	1,943,891	lbs/day

<u>OUT</u>			
12 Enzyme Recycle to Slurry Tanks	1,154,432	1,943,891	lbs/day

**FERMENTATION AND  
ENZYME, SUGAR, AND YEAST RECOVERY**

<u>IN</u>	<u>16 TON</u>	<u>27 TON</u>	
29 Liquids to Ultrafiltration	- 1,262,244 ✓	2,136,266 ✓	lbs/day
35 Dry Yeast	- 44 ✓	75 ✓	lbs/day
36 Ammonia	- 871 ✓	1,508 ✓	lbs/day
20 Fresh Water	- <u>1,577</u> ✓	<u>2,669</u> ✓	lbs/day
	1,264,736 ✓	2,140,518 ✓	lbs/day (2)

<u>OUT</u>	<u>16 TON</u>	<u>27 TON</u>	
14 UF Concentrate/ Enzyme Recycle	- 64,759 ✓	109,972 ✓	lbs/day
6 R. O. Water	- 1,110,422 ✓	1,875,462 ✓	lbs/day
37 Carbon Dioxide (CO <sub>2</sub> )	- 6,334 ✓	10,687 ✓	lbs/day
40 Yeast (70% Moisture)	- 1,054 ✓	1,784 ✓	lbs/day
42 Beer to Distillation	- <u>82,167</u> ✓	<u>142,613</u> ✓	lbs/day
	1,264,736 ✓	2,140,518 ✓	lbs/day (2)

# ETHANOL RECOVERY

IN	16 TON	27 TON	
42 Beer to Distillation	- 82,167 ✓	142,613 ✓	lbs/day
<i>Duplicate</i> { 3 Steam to Distillation	- 20,542 ✓	35,653 ✓	lbs/day
<i>nos</i> { 3 Steam to Pervaporation	- <u>493</u> ✓	<u>832</u> ✓	lbs/day
	103,202 ✓	179,098 ✓	lbs/day

OUT	16 TON	27 TON	
○ Fuel oil → <i>SS1 / 443</i>		<i>✓ ST3</i>	
44 200-Proof Ethanol	- 6,246 ✓	10,593 ✓	lbs/day
<i>word →</i> Still Bottoms	<i>75035</i> 75,592	<i>130745</i> 131,519	lbs/day
25 Evaporated Moisture	- 329 ✓	555 ✓	lbs/day
<i>Duplicate</i> { 45 Dist. Reboiler Condensate	- 20,542 ✓	35,653 ✓	lbs/day
<i>nos</i> { 45 Pervap. Reboiler Condensate	- <u>493</u> ✓	<u>832</u> ✓	lbs/day
	103,202 ✓	179,098 ✓	lbs/day
	<del>103,008</del>		
	103,202		
	<u>114</u>		
	0		

*Diff (Δ) →*

*Typo error*



OR 16T/27T DAY INPUT





**The University  
Of Alabama  
In Huntsville**

Huntsville, Alabama 35899  
Telephone: (205) 895-6361  
Fax: (205) 895-6668

Kenneth E. Johnson Research Center  
September 18, 1991

Mr. Michael C. Mensinger  
Manager, Process Development  
Institute of Gas Technology  
3424 South State Street  
Chicago, Illinois 60616

Dear Mr. Mensinger:

Over the past year since we participated in the Leo A. Daly Ethanol Production Facility feasibility study, there seems to have been a completely different group of people that surfaced in Washington on behalf of TVA seeking D.O.D. funding for a TVA acid hydrolysis demonstration plant. This group was not and had not been involved with any of the Redstone Arsenal plans prior to their discovery that we had already prepared the aforementioned feasibility study at the request of Redstone Arsenal officials. Since that time there has not to my knowledge been any communication between Titan, TVA, CERL, or IGT and the officials at Redstone regarding any competition or comparative study on our proposed enzymatic process and the TVA acid process.

I believe that the circumstances of this current evaluation by CERL and IGT of the Leo A. Daly study are totally unrelated to the purpose of the study that we prepared. I have attached the Scope of Work for the Leo A. Daly study for your information, and I hope that you will now understand some of my answers and comments relative to your inquiry. We were not even asked to provide a material balance for this study, but we felt compelled to include such information which was the basis for equipment sizing, capital cost estimate, and potential revenues.

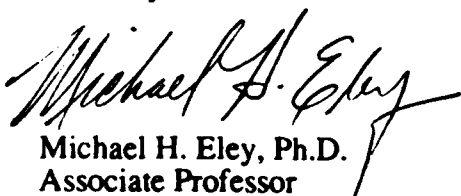
From my perspective, there has been no request from the people I have been working with at Redstone Arsenal to provide them with any additional information, but we have discussed how to proceed with the next step toward modification of their inactive incineration plant building and implementing a recycling project at Redstone. I am therefore confused about the purpose of your inquiry to make a comparison of enzymatic hydrolysis and the TVA acid hydrolysis processes with regard to the Redstone Arsenal study. Apparently, your inquiry is much more related to the TVA request for D.O.D. funding. Since I have just completed a 9-month sabbatical at TVA, I am aware that they are requesting a substantially higher amount of funding for their demonstration plant than is required for the small plant we proposed for Redstone Arsenal. I don't believe that the Leo A. Daly study was intended to be compared because it was very site specific, including using an existing building, limited feedstock, and some other very specific criteria that apply only to Redstone Arsenal. I feel that comparing enzymatic hydrolysis and acid hydrolysis under such circumstances puts the enzymatic process at an unfair disadvantage because of the limits of the Leo Daly study for Redstone Arsenal.

Since it would appear that we are not to be provided any other forum to present an appropriate comparison of the enzymatic hydrolysis process with the TVA acid hydrolysis process, I will attempt to respond to your questions and comments based on the Leo A. Daly study.

1. The information provided was sufficient to provide data to make the economic evaluation and was not intended to be an absolute material balance for final design and engineering. Thus, we did use certain numbers more than once, but did so because they came from the same source. For explanation, #3 was used wherever steam was needed because all steam is to be supplied by the Huntsville Solid Waste Disposal Authority steam plant to Redstone on a take or pay basis for the next 25 years. Stream #4 is a single source of hot water produced from various process steps and also used in several process steps. Stream #5 is still bottoms which goes through the hot water recycle (#4), but we continued to identify it as such because it is different in composition from other water recycle loops. Stream #6 is clean water from reverse osmosis, but it is used in several steps as make-up water. Stream #20 is fresh make-up water from the potable water system and is used in several cases where recycle water is insufficient due to evaporative losses in the process. Stream #25 is evaporated water where ever evaporation occurs. Stream #37 is CO<sub>2</sub> into and out of the recovery system. I don't know whether there are losses in recovery at this time, so it remained the same. Stream #45 is steam condensate that would be returned to the Huntsville steam plant. The steam for the fuel dryer should also be numbered #3 and the condensate from the dryer should be numbered #45 based on a later comment. We have no real problem with your comment, but we were not asked to provide a complete material balance other than to size equipment and determine economic feasibility.
2. Again, the study did not request an exact material balance. Fusel oil for example was only an estimate, since paper should have very little amino acids and fats, there may not be any fusel oil produced. We did not consider energy balance in the study. Steam was to be provided at no cost for our evaluation, and quantities of steam in closed loops that would be returned as condensate to the steam plant did not need to be absolutely correct. This is also the point of some of your later comments.
3. I failed to find any violations in the material balances we provided on the pages that you numbered 3 through 8. Your addition of steam to the material balance where the fuel dryer is used is unnecessary since that steam does not become part of the material being processed. Again, we were not asked for any energy balance data, particularly with steam usage and condensate return. The fusel oil was blank initially, but Daly added in a value later.
4. We did not include any value in our material balance for fusel oil, since there are no known amounts of amino acid or fats in paper, unlike grains, and thus there is no factual basis to assume any fusel oil at all. The Daly engineers felt it should be included, and an estimate was made by phone. I don't believe there will be anything close to 500 lbs per day of fusel oil in a paper only plant at Redstone in either size.
5. I have already explained the situation with steam in closed loops with condensate return to the steam plant. The numbers presented were essentially correct as far as process streams are concerned. The data presented were accurate enough for our study. I have also explained the fusel oil blank, and I see no need to labor on that point.

6. I have explained the steam and condensate return already concerning the fuel dryer. It was not necessary in our study because the steam was provided at no cost at the request of Redstone officials.
7. Other than the change made for fusel oil that may be incorrect anyway, there is a typo that you pointed out. If you make fusel oil zero as my balance did initially then the still bottoms are correct.
8. You are correct that the hammermilled paper remains the same and the amount of enzyme should also remain the same. However, since it all goes into hydrolysis, I believe the error is erased in the next step.
9. I believe I have already answered this question regarding steam source and its effects on our material balance and economics.
10. I agree that it will take somewhat more steam to evaporate water and these should not be equal, but again, it does not matter for our study since the cost of steam is zero and no energy balance was requested.
11. The composition of the two feedstocks are not known to us. We were told that stream #1 would be mixed, clean office paper, and the hammermilled paper would likely be the same, except it would be shredded. There has been no assessment made to my knowledge about quantities of various types of paper, such as computer print out, white ledger, etc.
12. As stated before, there was no request for energy balances, and steam was provided at no cost.
13. Some of these parameters are part of the proprietary information that was withheld as stated in our very brief statement in the forum of our study. I don't believe that the present circumstance of this inquiry alters the proprietary nature of such information.
14. The pervaporation system was a telephone quote, based on the beer concentration and throughput. Pervaporation was selected because it is less energy intense than distillation and does not require the use of potentially toxic or hazardous chemicals for azeotropic distillation.
15. This question also falls under our proprietary process information.

Sincerely,



Michael H. Eley, Ph.D.  
Associate Professor

attachment



21 May 90

Appendix A  
Scope of Work  
for

Ethanol Production Facility Study

Installation: Redstone Arsenal, Alabama.

A. GENERAL:

1. Description of the Project: Perform an Analysis of Building 5410, to determine the feasibility of converting this building into a ethanol production plant and preparing the DD Form 1391 to identify, analyze, and compare all reasonable alternatives available to Redstone Arsenal to utilize the waste paper produced at the installation.

a. Background This project is needed to develop an alternative energy source. Approximately 16 tons a day of paper is placed into landfills at the Arsenal. The revitalization of an existing refuse-fired power plant that will utilize clean waste paper, and through fermentation, produce ethanol is under consideration. The ethanol and waste by-product would be used as an off-setting fuel source.

2. Authorization: This project is authorized by an Installation Support Request, dated 16 May 90 from the Office of Directorate of Engineering and Housing (DEH), Redstone Arsenal, Alabama.

3. Coordination:

a. Point of Contact: Upon award of the delivery order, the A-E shall designate in writing a project coordinator or manager to serve as a single point of contact and liaison between the A-E and Contracting Officer and/or his representative for all services required by this delivery order.

b. Kansas City Project Manager: The Kansas City District (KCD) Project Manager for this project prior to delivery order award is Mr. Robert Miller, MRK-ED-MF, (816) 426-7348. The Project Manager is the primary point of contact for action concerning this delivery order prior to award.

c. Installation Points of Contact: After the delivery order is awarded it will be administered by Redstone Arsenal. The primary point of contact at Redstone Arsenal for this project is Mr. Sam Fields, Directorate of Engineering and Housing (DEH).

*Vickerman*

(205)-876-2126. Coordination of all facets of the project will be maintained with the Redstone Arsenal project manager. A minimum of fifteen (15) days notice shall be given by the A-E to the Redstone Arsenal Project Manager prior to any formal meeting at Redstone Arsenal. The A-E is required to coordinate all field activities with Mr. Fields prior to commencing those activities.

4. Public Disclosure: All data, reports, and materials contained or developed in this project are the property of the Government and will not be released by the A E or his consultants without written approval of the Contracting Officer.

5. Materials: All services, supplies, materials (except those specifically enumerated to be furnished by the Government), plant, labor, supervision and travel necessary to perform the work and render the data required under this contract are included in the lump sum price of the contract.

6. Records:

a. The A E shall provide a record of all conferences, meetings, discussions, verbal directions, telephone conversations, etc., with the Government representatives relative to this contract in which the A-E and/or designated representatives participated. These records shall be dated and shall identify the contract number, participating personnel, subject discussed, and conclusions reached. The A-E shall forward a copy of these records to the installation point of contact as soon as possible (not to exceed ten (10) calendar days) from the date of the event.

b. The A-E shall provide a record of requests for and/or receipt of Government-furnished material, supplies, data, documents, etc., which, if not furnished in a timely manner, would significantly impair the normal progression of work under this contract. The records shall be dated and identify the contract number, and the information requested or received. The A-E shall forward to the installation point of contact, as soon as possible (not to exceed ten (10) calendar days), a copy of the record of request or receipt.

c. The A-E shall submit a progress schedule before, or at the Kickoff meeting for approval by the Redstone Arsenal Contracting Officer. This schedule will include the dates for all definable work tasks, phases of work and significant events (milestones). The effort to complete each work task will be weighed against the total effort to complete the contract work. The percent complete of each definable work task will be indicated on the 15th of each month. The weighted effort times the percent complete of each task will be summed to determine the monthly percent complete for the entire project.

d. Provide to the Redstone Arsenal point of contact, copies of all miscellaneous data computations, references and studies used in developing this plan. The author, date and contract number must be on all miscellaneous data.

7. Submittal Format

a. Reports: The reports will be submitted in 8-1/2" x 11" format. All reports shall be spiral bound with the exception of 3 copies of the final report, which shall be submitted in hard cover 3-ring binders. Report text, charts, tables and graphs shall be Xerox (or equal) copies of the originals. A 3 1/2 double side, double density floppy disc of the final submittal of the narrative report shall also be submitted. The floppy disc shall be in Word Perfect, Macintosh compatible, computer software format.

b. Drawings: Diagrams and drawings will be submitted as required for clarity and as specified herein. Report covers and any report plans which are photo-reductions of plans shall be reproduced by offset printing medium.

c. Delivery: All submittals shall be delivered via an overnight delivery service directly to each reviewing agency listed in paragraph C., 3.

8. SERVICES TO BE PERFORMED BY ARCHITECT-ENGINEER (A-E):

1. Review Existing Documentation. The A-E shall review the following documentation:

A previous study of Building 5410..

2. Work Required.

a. Furnish expert review of the process of producing and utilizing ethanol.

b. Complete a Building Evaluation report of building 5410, an existing refuse-fired boiler plant. This plant was built to utilize waste paper but is not now in use.

c. Develop an economic analysis for a phased conversion of building 5410 to an operational ethanol production facility. The format shall be that of the ECON package on the DD 1391 Processor.

d. Evaluate five (5) existing power generation facilities and provide an estimate of capital improvement costs for the conversion of each facility to burn ethanol and anticipated by-products.

e. (Option): Develop a detailed DD Form 1391 with the 18 justification paragraphs to support the phased conversion and implementation of building 5410 to an ethanol production facility.

3. Interviews. The A-E shall interview representatives of the following organizations for input required for the study.

a. Redstone Arsenal's Directorate of Engineering and Housing (DEH), Planning and Construction Branch to gather information on existing power generation facilities.

b. Redstone Arsenal's Directorate of Engineering and Housing (DEH), Environmental Branch to gather information on wastes generated, environmental studies, and environmental regulations related to the power generation facilities.

c. There is a possibility of obtaining consulting help from the University of Alabama - Huntsville.

C. SUBMITTALS, MEETINGS AND REVIEWS:

1. Schedule:

Item	Days to Complete	Days from NTP
Kickoff Meeting	14	14
Valuation submittal	14	28
Building Evaluation Submittal	28	56
Draft DD Form 1391 submittal	28*	28
Completion of all work	14	42
* After acceptance by the Government of the Option and issuance of notice to proceed.		

2. Meetings: The A-E shall attend a preproposal meeting at the Kansas City District Office. Direct payment will not be made for this meeting. The A-E shall also attend the kickoff meeting listed in item C., 5. below. The meeting will be held at Redstone Arsenal. At least two (2) qualified representatives from the A-E's firm shall attend all meetings. One member of the original pre-negotiation team must attend all meetings.

3. Number of Copies per Submittal: For each of the submittals listed below, the A-E shall deliver three (3) copies of each required item directly to Redstone Arsenal except for the fourth submittal where five (5) copies shall be delivered to Redstone Arsenal and two (2) copies delivered to the Kansas City District.

4. Review Comments: The A E shall address and/or incorporate all review comments furnished by the Government. The A-E shall provide justification for not incorporating any review comment he feels is inappropriate.

**5. Project Submittals and Meetings:**

a. **Kickoff Meeting:** The meeting shall be held in the Redstone Arsenal DEH office. The general purpose of the meeting will be to allow the representatives of the various organizations to meet and discuss the scope of the contract work. During the meeting the A-E shall present his proposed schedule and method for accomplishing the contract work.

b. **Validation submittal:** The first submittal shall consist of three copies of the validation of the paper waste to ethanol process.

c. **Building evaluation submittal:** The second submittal shall consist of three copies of the draft building evaluation report for building 5410 and cost analysis of conversion of existing facilities.

d. **Third Submittal:** The third submittal shall consist of three copies of the draft DD 1391: Implementation of the Ethanol Production Facility Program.

e. **Fourth Submittal:** The fourth submittal shall consist of any corrections to the draft documents required by the Redstone Arsenal DEH and shall be furnished in five copies.

**6. Government Review of Work:**

a. Each report specified above will be reviewed by the Redstone Arsenal DEH to evaluate the adequacy of the work, the format of the documents, and the overall approach and direction of the project. The reviewers will indicate concurrence or nonconcurrence with the information developed and provided.

b. The Contracting Officer reserves the right to return for correction within the approximate review periods any and all products that are in error or, have not been prepared within the scope of work; unless otherwise notified in writing, these corrections will be incorporated in another work submittal. However, if any work submittal is in gross error, the Contracting Officer reserves the right to require a completely new submittal of the work.

**D. DATA TO BE FURNISHED BY THE GOVERNMENT:**

1. **Data to be furnished to the A-E:** The following information will be supplied to the A-E as data required for the study.

- a. A copy of the previous study.
- b. Drawings of building 5410.

## 2. REFERENCES:

### a. Army Publications:

- AR 200-2 Environmental Effects of Army Action
- TM 5-803-5. Installation Design, 1 Mar 81.
- TM 5-811-1 Electrical Power Supply and Distribution,  
12 Sep 84

## E. METHOD OF PAYMENT:

a. Title I Services The Architect-Engineer shall prepare and submit to the U.S. Army Engineer District, Kansas City, partial payment estimates using ENG Form 93, which shall serve as the request for payment. All partial payments shall be based on work completed as of the 15th day of the report month and shall be submitted to the office of the Contracting Officer by the 18th day of the month. The pay estimate shall be submitted with ENG Form 93, in accordance with the "Instructions for Completion of ENG Form 93 - Payment Estimate" dated 5 January 1983. The U.S. Army Engineer District, Kansas City, will prepare supporting payment documents after obtaining necessary approvals and forward all documents to the U.S. Army Engineer District, Omaha, for issuance of the payment check. All questions regarding payments shall be directed to the U.S. Army Engineer District, Kansas City.

F. ADDITIONAL CONFERENCES: Payment for furnishing the services of technically qualified representatives to attend conferences, other than those specified herein when so requested in writing by the Contracting Officer, will be made at the rate per hour for the discipline involved plus travel expenses computed in accordance with Government Joint Travel Regulations in effect at the time travel is performed and actual cost of transportation.

## G. ACCEPTANCE OF WORK:

1. Within 30 days after receipt of the Final Submittal, the A-E will be notified in writing of the acceptance of those documents.

2. Within 10 calendar days after notification that all work performed under the contract has been accepted, the A-E will deliver to the Contracting Officer the following:

a. All items and information furnished to the A-E by either the installation, or the Contracting Officer during the course of this contract.

b. All original text, maps, tapes, and other miscellaneous documents obtained or prepared under this contract.

## **DISTRIBUTION**

### **Chief of Engineers**

**ATTN: CEHEC-IM-LH (2)**

**ATTN: CEHEC-IM-LP (2)**

**ATTN: CERD-L**

**ATTN: CERD-ZA (5)**

**Fort Belvoir, VA 22060**

**ATTN: CECC-R**

**Defense Technical Info. Center 22304**

**ATTN: DTIC-FAB (2)**

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